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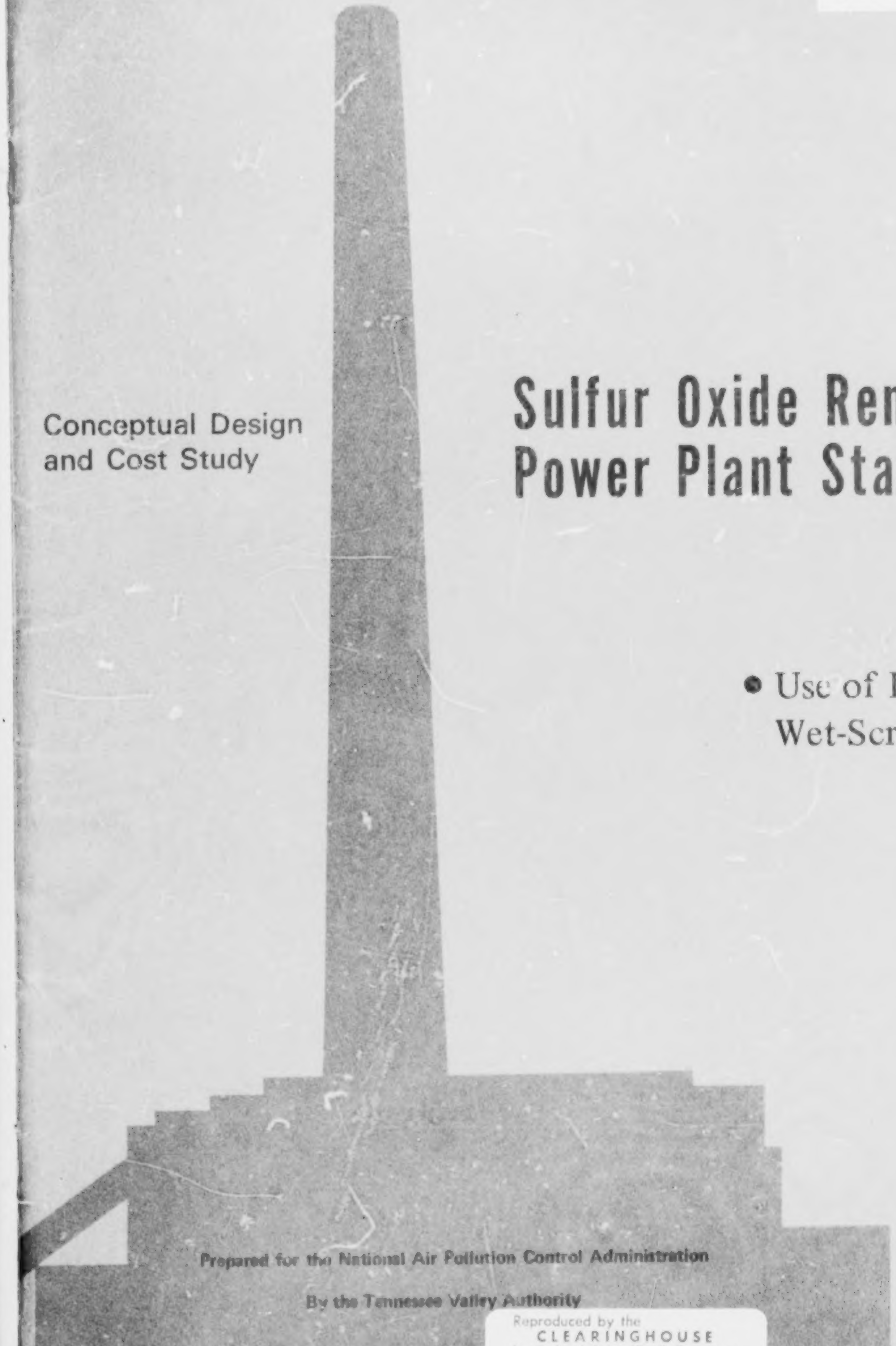

Conceptual Design
and Cost Study

Sulfur Oxide Removal From Power Plant Stack Gas

- Use of Limestone in Wet-Scrubbing Process

Prepared for the National Air Pollution Control Administration

By the Tennessee Valley Authority

1969

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Sulfur Oxide Removal From Power Plant Stack Gas

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Conceptual Design and Cost Study Series
Study No. 2

Prepared for
National Air Pollution Control Administration
(U. S. Department of Health, Education, and Welfare)
By
Tennessee Valley Authority

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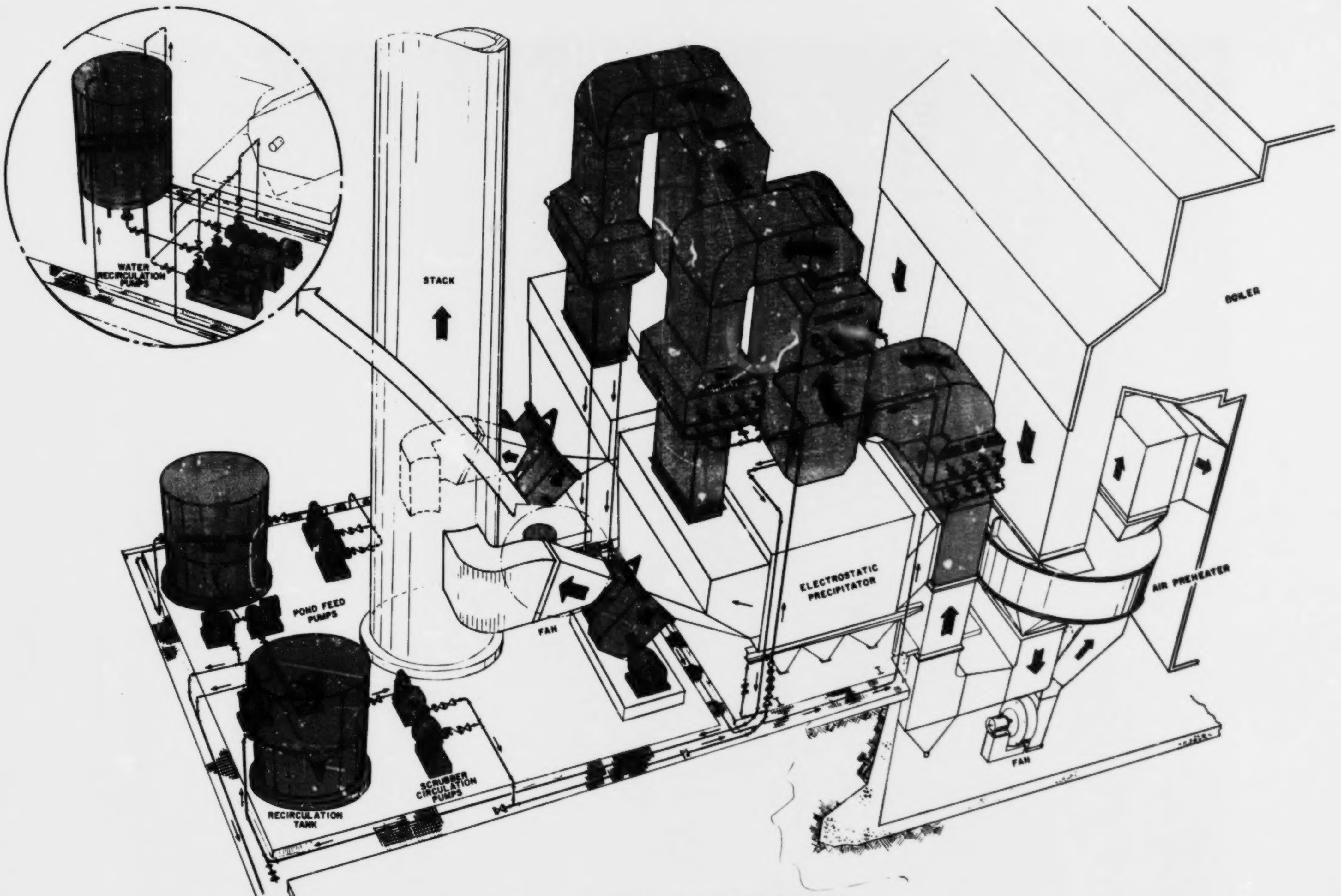
WATER RECIRCULATION PUMPS
STACK
BOILER
ELECTROSTATIC PRECIPITATOR
AIR PREHEATER
FAN
FAN
POND FEED PUMPS
SCRUBBER CIRCULATION PUMPS
RECIRCULATION TANK

POWER PLANT EQUIPPED WITH LIMESTONE–WET SCRUBBING PROCESS FOR SULFUR OXIDE REMOVAL

SCRUBBING EQUIPMENT SHOWN IN COLOR

Preface

On February 16, 1967, the National Center for Air Pollution Control (now National Air Pollution Control Administration), U. S. Department of Health, Education, and Welfare, entered into a contract with the Tennessee Valley Authority (TVA) for a series of conceptual design and economic studies to be carried out by TVA on processes for reducing sulfur oxide emissions from power generation. The purpose is to evaluate objectively and realistically the merits of different methods under consideration for sulfur oxide control, with a common and uniform basis used for comparison. The studies include (1) literature surveys, (2) direct contacts with workers in the field to supplement literature information, (3) specific studies by specialists in power plant design, chemical plant design, power plant operation, and pollutant dispersion to supplement the conceptual design study, (4) market surveys where appropriate, (5) quotations from vendors and fabricators on major equipment items, and finally (6) investment and operating cost estimates.

Work has proceeded on three processes: (1) limestone injection (dry process), (2) use of limestone in a wet-scrubbing process, and (3) ammonia scrubbing. The limestone injection study has been completed and a report issued: "Sulfur Oxide Removal from Power Plant Stack Gas–Sorption by Limestone or Lime (Dry Process)" (1968). The present report covers use of limestone in wet-scrubbing processes.

The work has been divided in TVA as follows:

Project Supervision

Applied Research Branch (Division of Chemical Development)

Cost Estimates

Applied Research Branch
Design Branch (Division of Chemical Development)

Design

Mechanical Design Branch (Division of Engineering Design
Design Branch

Report Preparation

Applied Research Branch
Research Staff (Office of Power)
Steam-Electric Generation Branch (Division of Power Production)
Air Quality Branch (Division of Health and Safety)
Water Quality Branch (Division of Health and Safety)

A major part of the evaluation has been the analysis of findings by other organizations who have worked on wet limestone scrubbing. These findings, which are, of course, the basis for the conceptual design, are used throughout the body of the report as needed. In addition, to serve as a general reference, the work of most of the organizations quoted is summarized and presented in appendixes. In addition to the published literature, several organizations have supplied information directly that has been quite helpful in the study; the contributions of the following are acknowledged.

Central Electricity Generating Board (England)
Combustion Engineering, Inc. - Detroit Edison Company
Imperial Chemical Industries Limited (England)
National Air Pollution Control Administration (NAPCA)
Union Electric Company
Universal Oil Products Company, Air Correction Division
Wisconsin Electric Power, Inc.

Copies of the reports in the NAPCA-TVA conceptual design series can be obtained by ordering from the following address.

Clearinghouse for Scientific
and Technical Information
5285 Port Royal Road
Springfield, Virginia 22151

The reports are identified and priced as follows:

Title	Number	Price
Sulfur Oxide Removal from Power Plant Stack Gas–Sorption by Limestone or Lime (Dry Process)	PB 178-972	$3.00
Sulfur Oxide Removal from Power Plant Stack Gas–Use of Limestone in Wet-Scrubbing Process	*	*

*Not yet assigned.

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Summary

The growing problem of atmospheric pollution by sulfur oxides has promoted a large amount of research and development on removal from power plant stack gases. Of the several processes that have been proposed, use of lime or limestone in a wet scrubber is one of the more promising. Although the sulfur oxides are recovered as calcium sulfate or sulfite, which are likely to be useless products in most instances, relative simplicity and low investment have promoted interest in the method. Moreover, the net cost penalty in producing power can be predicted with fair accuracy, whereas the net cost in recovery processes is uncertain because the market for the products is subject to variation.

Limestone–wet scrubbing can be used in two ways, either by injecting limestone into the power plant boiler and catching it in a wet scrubber after the air heater or by introducing limestone (or lime) directly into the scrubber system. Injection of limestone into the boiler removes a portion of the sulfur dioxide ahead of the scrubber, provides protection from corrosion by sulfur trioxide and alkali salts, and converts the limestone to quicklime, a more reactive form for use in the scrubber. Introduction of lime or limestone directly into the scrubber system eliminates potential boiler operating problems such as abnormal slagging and increased erosion and makes possible certain procedures that prevent equipment scaling. Direct introduction into the scrubber system was developed to a commercial scale in England over 30 years ago; modern emphasis is on the injection-scrubbing process.

The present study is one of a series of conceptual design and cost studies being carried out by the Tennessee Valley Authority (TVA) for the National Air Pollution Control Administration (NAPCA). The purpose is to develop, for various promising processes, the best design possible from existing data, estimate capital and operating costs on a uniform basis, and recommend further research and development needed. This is the second study in the series; the first was an evaluation of the dry limestone method–injection of limestone into the boiler without subsequent wet scrubbing. One of the objectives in the present study is to compare the merits of the wet and dry processes.

Other major work is being done in the area of limestone injection and wet scrubbing. In a joint program between TVA and NAPCA, various research projects now under way will culminate in large-scale field tests of the dry limestone process. The results of this study will be useful in refining evaluation of combined injection and wet scrubbing. Also, as part of the NAPCA program, Bechtel Corporation is preparing a preliminary design for a large pilot-scale study of limestone injection–wet scrubbing in scrubbers of various types.

In commercial application of the process, Combustion Engineering, Inc., is testing wet-scrubbing systems of the injection-scrubbing type at two power stations.

Process Problems

Results of experimental work demonstrate that a high degree (90% or more) of sulfur dioxide removal from power plant stack gases can be accomplished by scrubbing with a slurry of lime or limestone. Pilot plant results indicate that good utilization of absorbent can be expected when limestone is calcined by injection into the boiler or when lime is used directly. However, both full-scale and pilot plant tests show that absorbent requirement increases if uncalcined limestone (added in the scrubber system) is used as the absorbent. Parameters which affect degree of absorbent utilization–limestone type, particle size, scrubber operating conditions, and dead-burning (when limestone is injected in the boiler)–need further study.

Formation of sulfite and sulfate scale was reported to be a major problem in early commercial use of wet scrubbing. In the early work in England, a satisfactory method for preventing scaling was developed for the process in which lime or limestone is introduced directly into the scrubber system; the main feature was introduction of the absorbent into a delay tank after the scrubber. Similar scaling problems have been encountered in current plant tests in the country in which the limestone is injected into the boiler; prevention of scaling in such a system remains to be demonstrated.

Scrubbing with a water slurry cools the gases to the saturation temperature, about 125° F. Discharge of a cool saturated plume is not consistent with current emphasis on maximum plume rise for good dispersion. Also, there is some opposition to a visible vapor trail from a power plant stack. A study of the effect of reheat temperature on dispersion showed that with high scrubbing efficiency, reheat only to maintain temperatures above dew point to the stack top should prevent excessive ground-level concentration of sulfur dioxide. However, a higher degree of reheat may be required to prevent increase in ground concentration of other pollutants such as nitrogen oxides.

Disposal of large quantities of solid waste, 160,000 tons per year for a 200-megawatt plant, will require additional land area for settling ponds or for disposal in the solid form. Since the cost of recycling the sluice water is not high, recycling rather than release to streams is indicated. However, if high-calcium limestone is used as the absorbent, the product solids should not be soluble enough in the scrubber effluent to cause water quality problems in receiving streams of the size normally contiguous to power plants. If dolomitic limestone is used, the high solubility of magnesium salts would in many cases make sluice water recycling mandatory.

Study Assumptions

The cost of limestone–wet scrubbing and, to some extent, the type of equipment required, depend on several factors such as sulfur content of the coal, limestone cost, degree of grinding, power plant size and location, and excess limestone (above the theoretical amount) used to increase degree of sulfur oxide removal. It was necessary to assume a set of conditions as a base case for the conceptual design; in addition, the effect of varying each parameter was analyzed and factors developed wherever possible to relate the basic design and estimate to other conditions. The principal basic conditions assumed are given in the following tabulation:

Conditions Assumed for Study

Boiler size	200 megawatts
Fuel type	Pulverized coal
Sulfur content of coal	3.5%
Ash content of coal	12%
Plant location	Northwestern Alabama
Limestone type	High-calcium limestone ($94.9\%\ CaCO_3 + MgCO_3$)
Limestone cost	$2.05 per ton (delivered)
Particle size	70% -200 mesh
Amount of absorbent	110% of theoretical when injected into the boiler; 130% and 100% of theoretical, respectively, when limestone or lime is introduced into the scrubber system
Reheat temperature	250° F.

Cost comparisons made in the study include: existing versus new plants, operation with and without electrostatic precipitators in existing plants, alternate scrubber designs, alternate waste disposal methods, and alternate reheat methods and temperatures.

Process Equipment

The equipment needed for limestone injection followed by wet scrubbing includes many of the items required in the dry process. The main needs are a limestone-receiving hopper, storage facilities, a dry grinding installation, an injection system, a scrubber with mist eliminator, a gas reheat system, and solid waste disposal facilities. For limestone added directly in the scrubber system, a wet-grinding facility was provided.

For the absorption step, a dual-train, two-stage, mobile-bed scrubbing system was selected. The scrubber is designed to handle the total dust load. Gas reheat is effected by heat transfer from the gas ahead of the scrubber to the scrubber exit gas by circulating water through finned-tube exchangers.

Modification of the boiler induced-draft fans was necessary to compensate for the increased pressure drop (approximately 12 in. of water) in the system. Further investment is required for additional solids-disposal equipment and pond area to accommodate the increased waste load.

Costs

Capital and operating cost requirements have been calculated for such factors as unit size, sulfur content of coal, type of process, raw material cost, and operating time under the assumptions outlined above. The results are summarized in Tables 1 and 2.

Capital costs include all equipment from limestone

Table 1. Capital Requirements

Conditions	Capital, $/kw of power capacity
Base case[a]	13.05
Exceptions to base case	
2.0% S	11.70
5.0% S	14.30
Limited reheat[b]	
To 200° F.	10.52
To 175° F.	9.47
Process B[c]	13.80
Process B (with lime)[d]	20.00
500 mw	10.85
1,000 mw	8.21
1,000 mw, process B	8.82
1,000 mw, new power unit[e]	6.32

[a] Base case assumes 200-mw unit, existing power plant, 3.5% sulfur in coal, process A (injection-scrubbing), reheat to 250° F. by heat exchange, 99.5% dust removal, 95% SO_2 removal, and nonrecycle of sluice water.

[b] Reheat by direct firing natural gas.

[c] Addition of limestone to the scrubber circuit; 85% SO_2 removal.

[d] Addition of lime (CaO) to the scrubber circuit.

[e] Includes credit for eliminating electrostatic precipitator.

receiving to waste-solids disposal and stack gas reheat, plus installation, engineering and contractor fees, and provision for contingency. Since new units would not require the usual electrostatic precipitator, an investment credit equivalent to precipitator cost was taken. While there is some possibility that a shorter stack might be adequate for a new plant because most of the sulfur oxide is removed, other considerations such as particulate and nitrogen oxide emission as well as possible process upsets make this questionable; therefore no credit was taken for reduction in stack height.

Operating costs include raw material, labor, maintenance, and overhead costs; appropriate capital charges (including return on investment); assignable credits for eliminating precipitator usage (existing precipitators will not be operated) and corrosion reduction; and charges for thermal losses when limestone is injected into the boiler.

Table 2. Operating Cost

Conditions	$/ton of coal	Mills/kwh
Base case[a]	1.31	0.49
Exceptions to base case		
2.0% S	1.05	0.39
5.0% S	1.57	0.59
$1.00/ton limestone	1.17	0.44
$4.00/ton limestone	1.56	0.59
Intermittent operation[b]		
720 hr/yr (30 da)	0.75	0.28
4,000 hr/yr (½ yr)	1.18	0.44
Reduced boiler operation[c] (50% load factor)	2.10	0.79
Limited reheat		
To 200° F.	1.13	0.42
To 175° F.	1.06	0.40
Intermittent reheat to 200° F.[d] (30 da/yr)	1.07	0.40
Sluice water recycled	1.33	0.50
Process B[e]	1.42	0.53
Process B (with lime)[f]	1.85	0.69
500 mw	1.11	0.41
1,000 mw	0.90	0.34
1,000 mw, process B	0.99	0.37
1,000 mw, new power unit[g]	0.76	0.29
1,000 mw, new power unit, $1.00/ton limestone	0.64	0.24

[a] Base case assumes 200-mw unit, existing power plant, 3.5% sulfur in coal, 8,000 hr/yr operation, process A (injection-scrubbing), nonrecycle of sluice water, reheat to 250° F. by heat exchange, $2.05/ton limestone cost, 14.5% capital charge, 99.5% dust removal, and 95% SO_2 removal.

[b] Boiler operating at full load; scrubber operating for period shown; cost is average based on all coal burned and power produced/yr.

[c] Boiler operated intermittently or at reduced load; scrubber operated when boiler is in operation.

[d] Direct heating with natural gas.

[e] Addition of limestone to the scrubber circuit; 85% SO_2 removal.

[f] Addition of lime (CaO) to the scrubber circuit.

[g] Includes credit for eliminating electrostatic precipitator.

The cost of limestone is likely to vary widely, depending on amount used, power plant location, and the shipping mode and distance. In a survey of the limestone industry presented in this study, it was indicated that the price range will be roughly $1.00 to $3.00 per ton (delivered, -1/2-in. size). A very large amount of limestone is required if the process is operated continuously—on the order of 450,000 tons per year for a 1000-megawatt plant. Under conditions of large-scale usage, it may be possible to get special price and shipping cost considerations.

Conclusions and Recommendations

Limestone–wet scrubbing has some important advantages over the dry limestone process, including better efficiency for removing sulfur oxides (over 90% vs. 50-60%), less potential interference with boiler operation, and generally lower operating cost for large power plants. On the other hand, the dry process releases the stack gas at higher temperature (thereby giving better dispersion of residual pollutants), requires less investment, is simpler to operate, is more amenable to intermittent operation, and is less likely to cause water pollution. The wet process appears to be the best choice for the larger power plants, particularly those on baseload service. Moreover, if potential problems can be worked out, the wet process may have an economic advantage over the dry method in small plants, for comparable degrees of sulfur oxide removal. The dry method is more applicable to situations where a high degree of sulfur oxide removal is not required and where intermittent operation to reduce cost is feasible or the boiler is a peak-load unit out of service from time to time. The relatively low investment for the dry process reduces the penalty of continuing capital charges when the unit is out of operation.

Of the two wet-process variations, introduction of lime or limestone into the scrubber system costs somewhat more; lime is too expensive to consider and more absorbent is required when limestone is used. However, the limestone handicap may be removed by further research. The method is attractive because of its proven effectiveness in eliminating scaling.

Research and development should be pushed forward vigorously on the limestone–wet scrubbing process because of its promise as a basic method for preventing pollution by sulfur oxides. The main areas needing investigation are (1) the basic chemistry of the process, (2) methods for preventing scaling, including further development of adding limestone in the scrubber system, and (3) testing of equipment arrangements for intermittent operation and for partial sulfur oxide removal.

Introduction

This report presents the second of a series of design and cost studies made by TVA (Tennessee Valley Authority) for NAPCA (National Air Pollution Control Administration) on methods for removing sulfur oxides from power plant stack gases. The first covered sorption by limestone[1] in a dry method (1); the present one is concerned with use of limestone in a wet-scrubbing process.

Of the various atmospheric pollutants, sulfur dioxide is generally regarded as one of the most serious. Approximately 30 million tons of sulfur dioxide is emitted to the atmosphere annually in the United States, and power plants contribute about half. Hence the problem of reducing sulfur dioxide emission from power plants is receiving increased emphasis. Several ways of reducing emission have been studied, including use of low-sulfur fuel, fuel desulfurization, fuel gasification to give a clean fuel to the boiler, and sulfur oxide removal from the stack gas. Of these, stack gas treatment has received major attention because of its several advantages and because it has the greatest potential for near-future application.

Stack gas treatment processes that have been studied fall into two classes–those in which the sulfur is recovered in a useful and salable form and those of the nonrecovery ("throwaway") type. The throwaway methods, while giving no return to help offset operating cost, have the advantages of relative process simplicity and of avoiding the complexities in chemical product marketing. The sulfur oxides are sorbed by lime or limestone–the only sorbent low enough in cost to consider–and the resulting calcium sulfite or sulfate discarded.

Limestone sorption processes are of two types–wet and dry. Dry sorption involves injection of finely divided lime or limestone into the boiler and collection of the calcium sulfate product, along with the fly ash, in the usual dust-removal equipment after the boiler. In the wet process, a slurry of lime or limestone absorbs the sulfur oxides in a scrubber at the end of the system. The absorbent can be introduced directly into the scrubber recirculating slurry or it can be injected into the boiler as in the dry process.[2] The main advantage of the latter is that the injected limestone is calcined in the boiler and the resulting lime (which is caught in the scrubber) is a better absorbent than limestone; calcination in separate equipment would be quite expensive.

The main advantage of the wet process over the dry method is more complete utilization of the lime or limestone. The longer retention time afforded by the wet scrubber, plus the more favorable conditions for mass transfer, give on the order of three to fivefold more complete reaction of the absorbent. A further advantage is that the wet scrubber can remove the fly ash as well as the sulfur oxides. But offsetting these good points are the higher investment for the wet process and the loss in stack gas buoyancy resulting from cooling in the scrubber. Some degree of reheating is generally considered necessary, even with a high degree of sulfur oxide removal, to improve dispersion of other pollutants such as nitrogen oxides and residual dust. The economic effects of these opposing factors have been evaluated and are presented in this report.

Much of the background for use of limestone as a sulfur oxide sorbent, e.g., limestone availability and mining technology, was given in the earlier report,[3] and has not been repeated in the present one. It will be helpful to those using the present report to have a copy of the previous one available.

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[1] Throughout this report the word "limestone" is used in a general sense to cover all limestones and dolomites, including high-calcium limestone (mainly $CaCO_3$), dolomite ($CaCO_3 \cdot MgCO_3$), and compositions that give intermediate CaO:MgO ratios.

[2] The following "names" are used to distinguish between the three process types and combinations thereof:

Dry limestone injection: No scrubber involved

Limestone–wet scrubbing: General term to distinguish from dry injection; no specification as to point of limestone addition.

Limestone injection-scrubbing: Injection into boiler followed by wet scrubbing.

Limestone–scrubber addition: Introduction into scrubber recirculating liquor.

[3] See preface page 3 for title and ordering information.

History and Status

The use of limestone or lime as an absorbent for sulfur oxides in a wet process has been studied by several research organizations over a period of about 30 years. Most of the work has been small-scale, but some pilot plant and plant tests have been made. In this section of the report the previous work will be summarized as background for the design and cost study.

As noted earlier, two methods for addition of limestone to the system have been studied. Introduction of the reactant directly into the scrubber system was studied first, dating back to the late 1920's. Research programs have been carried out by the British Electricity Authority and the London Power Company (2, 3), Imperial Chemical Industries Limited - James Howden and Co., Ltd. (4), TVA (5), the Scientific Research Institute (Soviet Union) (6), and Wisconsin Electric Power Company (7). The two steps in limestone injection-scrubbing have been studied both separately and together. Research on the injection step (without scrubbing) has been conducted by many organizations, particularly in Germany from 1950 to the present; the history of this work is detailed in the dry limestone injection report (1). In addition, TVA and NAPCA are now studying the process in a program involving a full-scale boiler. Both steps, injection and scrubbing, were first studied together by Combustion Engineering, Inc. (8).

All of these studies were concerned with removal of sulfur oxides from power plant gases. In addition, Murakami and Hori (9) have developed a process for scrubbing tail gas from sulfuric acid plants and other chemical operations in which a lime slurry is used as the absorbent.

Summaries of most of the studies listed above are given in Appendix A.

The London Power Company did pilot plant work that led to full-scale gas-washing plants at the Battersea and Bankside power stations in London. Alkaline water (alkalinity equivalent to 200 p.p.m. of sodium hydroxide) from the Thames River was used for scrubbing the gas on a once-through basis. During part of the program lime or calcium carbonate (chalk) was added for pH control. The system removed a high percentage of the sulfur dioxide (about 90%) but the acidic effluent from the two stations lowered the pH of the Thames to an undesirably low level and thus prevented use of the process at additional plants. The once-through system also had the drawbacks of requiring a very large amount of water and of cooling the gas to an unduly low temperature.

In another study, a process utilizing a circulating liquor, kept alkaline by addition of lime or chalk, was developed by James Howden and Co., Ltd., and Imperial Chemical Industries Limited. The process was used commercially at the Tir John (Swansea) and Fulham (London) power stations. In these plants, both lime and finely ground chalk were tested; lime reacted stoichiometrically with sulfur dioxide but about 30% excess calcium carbonate was required for comparable sulfur dioxide removal, presumably because of slower reaction rate. About 98% removal of both dust and sulfur dioxide was achieved. Scaling was controlled but corrosion was a problem.

Little further work has been reported in England in the past three decades. The Tir John plant was not operated very long and the Fulham plant was shut down during World War II for military reasons (because the plume was believed to be guiding enemy aircraft); it was not rehabilitated after the war (2) because of the cost and because emphasis in sulfur oxide control had shifted to use of high stacks. As far as is known, the Battersea and Bankside once-through scrubbers are still in operation.

During brief pilot plant studies by TVA (5) in the early 1950's, a packed tower was used to scrub stack gas with a 10% slurry of pulverized limestone (95% -100 mesh). About 89% sulfur dioxide removal was obtained with stoichiometric addition. Evaporative cooling of the inlet gas was necessary for good scrubber performance.

Pilot plant scrubber studies were made in 1965 in a joint program between Wisconsin Electric Power and Universal Oil Products, Air Correction Division (7). A mobile-bed scrubber (Turbulent Contact Absorber), designed for operation at a gas flow rate of about 2000 cubic feet per minute, was used to scrub a portion of the gas from a 120-megawatt pulverized-fuel boiler at Wisconsin Electric's Oak Creek station. Circulating a slurry of pulverized limestone (90% -200 mesh) through the scrubber removed about 85% of the sulfur dioxide and 99% of the fly ash. In a separate study, dry injection of limestone was investigated but the combination of dry injection followed by wet scrubbing was not tested.

Murakami and Hori (9) developed a process for removing low concentrations of sulfur dioxide from industrial waste gases. The gases are cooled in a spray chamber where dust is removed and then passed through two scrubbers in series where sulfur dioxide is absorbed in a slurry of lime and calcium sulfite. The calcium sulfite is oxidized with atomized air and gypsum is separated by centrifuging; the liquid phase is recirculated to the absorbers. High-quality gypsum is produced. The method is reported to be in use in two plants in Japan, one a sulfuric acid plant and the other a chemical plant in which gases from an oil-fired tunnel kiln are scrubbed. Mitsubishi Heavy Industries, Ltd., is offering the process for use in power plants (10).

All of these studies were concerned with the scrubbing step only and demonstrated that lime is a good absorbent, capable of removing 90% or more of the sulfur oxides. Raw limestone under similar scrubbing conditions gave somewhat lower efficiency.

The concept of calcining the lime prior to scrubbing by injecting it into the boiler apparently was first proposed by Wisconsin Electric but was first tested by Combustion Engineering and Detroit Edison in 1966-67 (8). In the initial studies, with a pilot plant designed to scrub about 1000 cubic feet per minute, mixtures of dolomite and fly ash were injected into the stack gas ahead of the scrubber. When calcined dolomite was added at a rate equivalent to 120% of stoichiometric, 99% sulfur dioxide removal was achieved; use of raw dolomite gave only about half this efficiency. Dust removal of 98 to 99.6% was obtained. These encouraging results led to the installation of a 3000-cubic-foot-per-minute recirculating scrubbing system at the St. Clair station of the Detroit Edison Company. Dolomite was injected into the boiler and collected in a mobile-bed scrubber. At 110% stoichiometric dolomite addition, the system removed 98% of the sulfur dioxide and 99.5% of the dust.

On the basis of the pilot-scale work, Combustion Engineering now offers the injection-scrubbing process on a guaranteed basis and systems have been purchased by the Union Electric Company (St. Louis, Missouri) and the Kansas Power and Light Company (Lawrence, Kansas). A 140-megawatt tangentially fired boiler at Union Electric's Meramec station in St. Louis has been equipped with facilities for separate injection of limestone through the top row of burners, scrubbing the gas in a mobile-bed scrubber (National Dust Collector Hydro-Filter), partial reheating of the gas, and disposing of waste solids. The scrubber was designed for removal of at least 82% of the sulfur dioxide and 99% of the dust. Existing electrostatic precipitators will not be used; gas flow is diverted through by-pass duct work to two parallel scrubbers. The flow to each scrubber is further split into 40% and 60% proportions to allow for turndown during operation at reduced loads.

Operation of the Union Electric system was started in late 1968. Results of the preliminary operation have been reported by Jonakin (11). Scaling and plugging of the scrubber, particularly in the lower part, has been a severe problem. Sulfur oxide removal has also been below the design figure, but this may be due to poor gas and liquor distribution resulting from the scaling. Various procedures are being tested in an effort to avoid the problem.

In other announced applications of this process, Kansas Power and Light is currently equipping an existing 125-megawatt unit at the Lawrence station with scrubbing facilities provided by Combustion Engineering. In this installation, limestone and coal will be pulverized together in existing mills and injected through all burners. A new 420-megawatt unit scheduled for operation in 1970 will also be equipped with a limestone injection-scrubbing system.

The history of limestone–wet scrubbing development indicates that the method is a good one but that several problems remain, the most serious being scaling of equipment. Introduction of lime or limestone into the scrubber circuit rather than into the boiler may be a solution to the problem, as the Howden-ICI group were able to operate in this way without scaling. However, lime is quite expensive and limestone is not as efficient as in boiler injection. Work is needed on increasing limestone utilization.

Process Chemistry and Kinetics

Very little in the way of chemical and kinetic data has been reported on the limestone–wet scrubbing system. Although several organizations have done applied research, apparently only the Howden-ICI group (4, 12, 13) and Chertkov and coworkers (6) in the USSR have reported any basic data on the system. Data are available in the literature on the system $CaO\text{-}SO_2\text{-}H_2O$ but these are not considered applicable because of the absence of carbon dioxide.

One of the major questions is the nature of the calcium compound that actually reacts with the sulfur dioxide, which obviously is important in regard to reactant utilization and process equipment requirements. Pearson et al. (4), in their report of the Howden-ICI work, state that the effective reactant is calcium bicarbonate, formed by reaction of calcium oxide or calcium carbonate with dissolved carbon dioxide. This would be expected since calcium bicarbonate is reported to be relatively soluble as compared with the other calcium compounds present. Moreover, the large partial pressure of carbon dioxide promotes conversion of any dissolved lime to carbonate or bicarbonate.

Pearson, et al. postulated the following reactions:

$$CaO + H_2O \rightarrow Ca(OH)_2 \tag{1}$$

$$Ca(OH)_2 + CO_2 \rightarrow CaCO_3 + H_2O \tag{2}$$

$$CaCO_3 + CO_2 + H_2O \rightarrow Ca(HCO_3)_2 \tag{3}$$

$$Ca(HCO_3)_2 + SO_2 + H_2O \rightarrow CaSO_3 \cdot 2H_2O \downarrow + 2CO_2 \tag{4}$$

$$CaSO_3 \cdot 2H_2O + \tfrac{1}{2}O_2 \rightarrow CaSO_4 \cdot 2H_2O \downarrow \tag{5}$$

Chertkov (14) supports the conclusion that calcium bicarbonate is the effective reactant and gives data on the system.

It can also be postulated that hydrated lime reacts directly with sulfur dioxide, a reaction that is known to occur in scrubbing sulfuric acid plant tail gas with lime slurry (in the absence of carbon dioxide). No data appear to be available on the competition between this reaction and those listed above.

Or the system could be similar to ammonia scrubbing, where it is generally considered that carbonate is not involved and that sulfite is the principal reactant.

$$(NH_4)_2SO_3 + SO_2 + H_2O \rightarrow 2NH_4HSO_3 \tag{6}$$

Ammonia added to the recirculating scrubber then regenerates the sulfite.

$$NH_4HSO_3 + NH_3 \rightarrow (NH_4)_2SO_3 \tag{7}$$

Such a system seems unlikely in lime scrubbing, however, as calcium sulfite is quite insoluble and is the product formed–and precipitated–instead of calcium bisulfite.

Lessing (13) reported analyses of the circulating liquor in the Fulham pilot plant (Howden-ICI process). The range for several samples taken during one run, at points between the scrubber and the lime mixing tank, was (for the filtered liquor):

	Parts/million
$Ca(HSO_3)_2$	Nil
$CaSO_3$	10-40
$CaSO_4$	2170-2920
$CaCO_3$	150-220
$Ca(OH)_2$	Nil

The significance of this analysis is not clear. The author points out that the samples were not analyzed immediately, so oxidation of sulfite may have taken place. Also, it is not clear whether the analytical method would distinguish between $CaCO_3$ and $Ca(HCO_3)_2$. The very low CO_3^{2-} content casts some doubt on the presence of soluble $Ca(HCO_3)_2$; however, since the sample was taken after the scrubber and before lime addition, dissolved bicarbonate (or hydroxide) may have already reacted with sulfur dioxide.

The chemistry of limestone–wet scrubbing is complicated by the fact that impurities can contribute other products and the magnesium present, if dolomite or dolomitic limestone is used, produces magnesium sulfite and sulfate. In contrast to the other sulfur compounds formed, magnesium sulfate is quite soluble and therefore would increase the dissolved solids content of the slurry returning to the scrubber and going to the disposal pond.

Thus the scrubber slurry contains mainly $CaCO_3$, $CaSO_3$, and $CaSO_4$ (plus $MgSO_3$ when dolomite is used) in suspension and a variety of compounds and ions in solution. The solubility relationships in the complex phase system resulting have not been worked out. Further investigation is highly desirable, to give a better understanding both of sulfur dioxide absorption and of scrubber scaling caused by deposition of precipitating solids.

Engineering considerations require that the absorbent slurry be recirculated to the scrubber, with a sidestream drawn off to remove the "make" of calcium sulfite and sulfate. Thus the recirculating slurry changes composition at two points, where the absorbent is added and where the stack gas contacts the slurry in the scrubber. The sulfur dioxide absorption reduces the pH, but not radically because, according to Pearson, et al., the bicarbonate reacting with the sulfur dioxide is replaced by reaction of suspended calcium carbonate with carbon dioxide, thus giving a stabilizing effect on pH (4). At the point of lime or limestone addition, however, pH does tend to increase, particularly when the absorbent is added after the scrubber. The main effect of this, as reported by Nonhebel (12), is to force calcium sulfite and sulfate out of solution; for the pH increase involved, from 6.2 to 6.8, solubility of calcium sulfite decreased by about 20%. No data were presented for calcium sulfate, but it seems likely that some sulfate precipitation would take place also. More data are needed relating pH in the system to composition of the liquid phase of the slurry. This is important in relation to the scaling problem, as is the degree of supersaturation to be expected at various points in the scrubber circuit.

Oxidation of sulfite is another phase of the process chemistry, and also one that is not well understood. It has been generally considered that oxidation is desirable, because if the sulfite is allowed to escape into watercourses it will gradually oxidize there and reduce the oxygen content of the water. Moreover, in the Howden-ICI tests sulfite was said to interfere with prevention of sulfate scaling. Oxidation before disposal was a major objective in the British work.

Lessing (13) found that sulfite was rapidly oxidized in solution but not as a solid after precipitation. Solubility of sulfite increases as pH decreases; by keeping the pH below 6.4 at the scrubber outlet, the make of sulfite was said to be kept in solution. However, scrubbing efficiency drops with pH, so good control to hold the pH below and as close as possible to 6.4 was necessary. Although there is a hundredfold or so excess of oxygen in the stack gas (over the amount required for SO_2 oxidation), only about 50% oxidation was achieved in the Howden-ICI pilot plant scrubber. When the scrubber liquor was passed through a separate tower in countercurrent flow with air (about 2% of the combustion gases), nearly complete oxidation was obtained. It was speculated that small amounts of manganese from the coal and limestone served as an oxidation catalyst. In the earlier work at Battersea, aeration of the effluent in the presence of iron reduced sulfite concentration by a factor of 10.

In the pilot plant work by Combustion Engineering (8), the scrubbing effluent contained about equal amounts of sulfate and sulfite. Similar results were obtained in the Howden-ICI work (4). Murakami and Hori (9) use a pressure oxidation step to convert sulfite to sulfate, the objective being to make salable calcium sulfate.

During studies of absorption of sulfur dioxide in lime slurries, Chertkov (15) concluded that the rate of oxidation of sulfite to sulfate was proportional to the amount of calcium bisulfite in the scrubber liquor. Addition of a small amount (0.002-0.004%) of p-aminophenol reduced oxidation three to fivefold.

Other questions can be raised regarding oxidation of sulfite–effect of oxidation on scrubbing efficiency, relative oxidation rate of calcium and magnesium sulfites, further ways to inhibit or promote oxidation (analogous to data obtained in ammonia scrubbing), and desirability of oxidizing relatively insoluble magnesium sulfite to soluble magnesium sulfate. Few data are available that would help in answering any of these questions.

Little information is available also on the kinetics of the various reactions and transfers that take place in the scrubber–hydration of calcium oxide and dissolution of calcium hydroxide in the boiler injection process, dissolution of limestone when the limestone is introduced into the recirculating liquor, possibly the absorption of carbon dioxide in the scrubber, and, of course, transfer of sulfur dioxide to the reaction site and reaction with the absorbing species, whatever it is. Data on these must be obtained before the controlling mechanism or mechanisms can be determined.

For transfer of the gaseous solutes, a molecule of solute must be moved from the main body of gas to the gas-liquid interface, absorbed into the liquid (with or without chemical reaction), and moved from the interface into the liquid by diffusion or turbulence. Degree and rate of absorption depend on the difference between the partial pressure of the gaseous solute and its vapor pressure above the absorbing liquor. Since the concentration of sulfur dioxide in power plant stack gas is low (0.1-0.3%), the vapor pressure above the scrubbing solution must be quite low to provide a driving force for absorption. A complicating factor is that the temperature of the scrubbing liquor governs the vapor pressure of absorbed sulfur dioxide at a given concentration, and, without external cooling, this liquor temperature is fixed at about 125° F. by the heat content (wet bulb temperature) of the stack gas. Moreover, low pH gives high vapor pressure when lime is used as the absorbent; the effect of pH on vapor pressure of sulfur dioxide above calcium sulfite-bisulfite solutions, as determined by Chertkov (6), is shown in Figure 1. Therefore, the vapor pressure must be kept low by maintaining the effective concentration of dissolved sulfur dioxide at a low level, either by scrubbing with a very large amount of water as was done at Battersea and Bankside or by keeping the pH high by lime addition as was done in the cyclic processes.

Presumably, mass transfer of carbon dioxide is not a problem since its concentration in the gas stream is relatively high; however, no data are available on the transfer rate.

Transfer rate of the calcium may be an important consideration, especially when raw limestone or dolomite is used. Here the solute molecule does not have to move through a layer of molecules to the interface, as in transfer of gases, but must be pulled away from neighboring

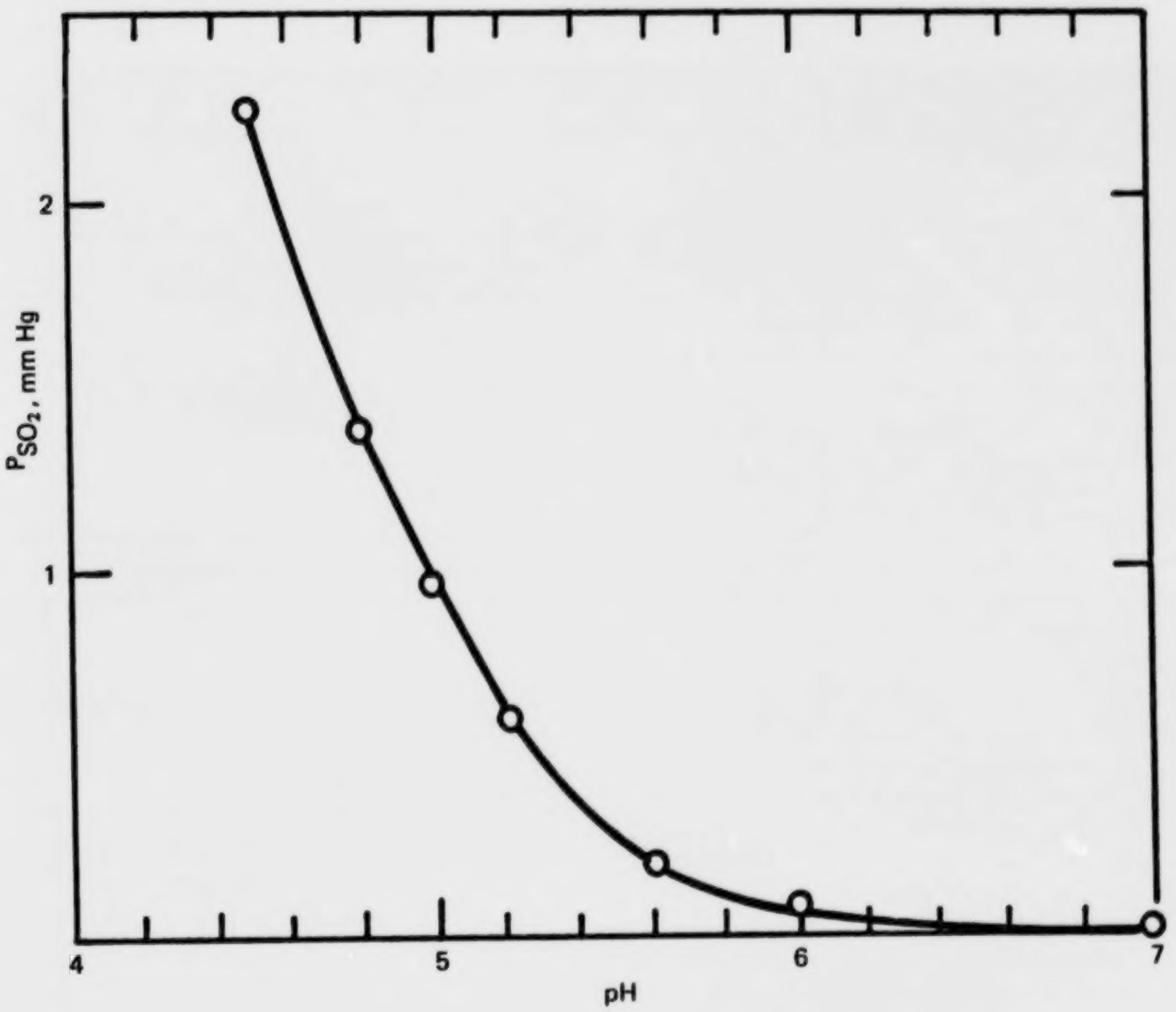

Figure 1. Effect of pH of calcium sulfite-bisulfite solution on SO_2 equilibrium vapor pressure

molecules to which it is much more tightly held than in a gas mixture. Thus particle size, degree of crystallinity, crystallite size, deposition of reaction products on particle surfaces, and other factors enter. The kinetics in such a stiuation are somewhat complicated and have not been studied for the CaO (or MgO)-CO_2-SO_2-H_2O system.

Available data indicate that lime (CaO) does better than raw limestone, presumably because the small particle size resulting from calcination and subsequent hydration makes it dissolve faster. Adequate data are not available, however, to evaluate either this factor or other parameters that may affect rate of dissolution.

Some data have been obtained on overall transfer rates in pilot plant scrubbers. Chertkov (6) reported that K (lb./sq. ft.-hr.-atm.) increases in direct proportion to the gas flow rate at low values of the latter. The increase in K was from 40 to 80 as the gas velocity was increased from 1 to 2.7 feet per second in scrubbing gases containing 0.2 to 0.3% sulfur dioxide in a grid-packed absorber with lime slurry containing 32% solids. The coefficient also increased with decreasing solids content in the scrubbing slurry; with 20% solids, the value of K was 160 at a gas velocity of 2.7 feet per second.

In summary, the chemistry and kinetics of the limestone–wet scrubbing process are quite complex and largely unexplored. Most of the investigations have been of the applied type, with the result that the basic mechanisms and reaction kinetics have not been studied adequately. Data are needed especially on (1) the main reaction mechanisms involved, (2) the rates of these reactions, (3) the transfer rates of reactants, and (4) the solubility relationships in the scrubber solution.

Process Considerations and Problems

The preceding history of the limestone–wet scrubbing process and the discussions of process chemistry and kinetics leave several questions that must be resolved before settling on a conceptual design. There are so many considerations that it is difficult to define a basic process that gives the best combination of the various parameters. The considerations that bear directly on process selection will be examined first.

Point of Limestone Addition

The major process choice to be made is between injecting limestone into the boiler and introducing lime or limestone into the scrubber circuit. The early workers apparently never considered the possible advantages of injecting into the boiler and worked rather on the problems of introducing the absorbent into the recirculating scrubber liquor. The practice has grown, however, of injecting small amounts of a basic material, usually calcium or magnesium oxide or carbonate, into boilers–not to remove sulfur dioxide but rather to reduce boiler corrosion. Thus when attention was turned again in recent years to reducing pollution caused by sulfur oxide emission, injection of limestone into the boiler as a means of controlling both pollution and corrosion gained favor. But although this appears quite attractive, there are major potential drawbacks that may offset the advantages. The large-scale tests now under way at the Union Electric and Kansas Power and Light plants should provide useful data, as will the forthcoming tests of dry limestone injection by TVA (1). However, since no large-scale tests of scrubbing with absorbent introduced directly into the scrubber circuit are planned, the Howden-ICI experience of 35 years ago must be used (along with TVA unpublished data) for comparison.

The good points of limestone injection-scrubbing can be summarized as follows:

1. *Better absorbent utilization.* Limestone injected into the combustion space of a boiler is calcined almost instantaneously. This is a prerequisite for the dry limestone injection method and is also helpful in the wet process, because the freshly calcined lime passing directly to the scrubber at the end of the system goes into solution more rapidly than does raw limestone. As noted earlier, those investigators who have compared calcined and raw limestone all found that more absorbent was required when uncalcined limestone was used.

The limestone does not have to be calcined in the boiler, of course; in the British work, lime made in the usual manner, in commercial limekilns, was used. However, cost estimates presented in the report of the TVA dry limestone injection study showed that calcining in the boiler is an economical procedure and that the cost of calcining in separate facilities is prohibitive.

Hence economics would dictate use of raw limestone if the absorbent is introduced into the scrubber circuit rather than into the boiler. On the basis of past work, it must be assumed that more absorbent would be required by this procedure and that injection into the boiler would therefore have a major advantage. However, there does not appear to be any reason for raw limestone to be inferior if it can be dissolved at an adequate rate; further development work may make this possible.

2. *Reduced boiler corrosion.* The potential reduction of both high- and low-temperature corrosion in the boiler system as a result of limestone injection was discussed in the dry process report. The tests at Detroit Edison indicated that reaction of the limestone with corrosive alkali salts, which normally reduce tube life in the areas exposed to maximum heat, should reduce high-temperature corrosion. Also, virtual elimination of sulfur trioxide in the combustion gas should decrease any corrosion in the air heaters caused by condensation of sulfuric acid.

However, the stoichiometric amount of absorbent needed for sulfur dioxide removal is far more than needed to react with the relatively small amount of alkali salts and sulfur trioxide present in the gas stream. It should be possible to avoid the potential drawbacks of boiler injection and still retain the advantage of corrosion reduction by injecting only a small amount into the boiler and adding the remainder in the scrubber circuit. Some additional investment would be required but it should not be large.

Injection of part or all of the limestone into the boiler would have a major advantage in using one of the more promising methods for stack gas reheating, that of exchanging heat between gas before and after the scrubber. Without limestone injection into the boiler, the sulfur trioxide would condense in the form of sulfuric acid as the gas cooled in the exchanger, and it would be necessary therefore to use relatively expensive construction material.

3. *Sulfur dioxide sorption in the boiler.* The limestone injected into the boiler should react with part

of the sulfur dioxide as it does in the dry injection method. However, since the amount of limestone needed for good overall sulfur dioxide removal in the wet process is considered to be near stoichiometric, the degree of sorption in the boiler would not be expected to be more than 20 to 30% because of the intrinsic inefficiency of the dry process. Getting this much reaction in the boiler is desirable, since it does not have to be done in the scrubber. The net effect is that the gas entering the scrubber would contain essentially no sulfur trioxide and less sulfur dioxide. This might reduce the scaling problem, as the calcium sulfate formed in the boiler would not have to be precipitated in the scrubber. Scrubber size and circulation rates, however, probably would not be benefitted.

It should be noted that even 20 to 30% sulfur dioxide removal in the dry injection process is expected to require very careful operation in regard to point, angle, and velocity of injection and possibly very fine particle size. Going to all this trouble may not be warranted in the wet process because any sulfur dioxide not reacted in the boiler will be caught in the scrubber. Hence the degree of reaction in the boiler may be quite low in practice.

Offsetting these advantages are the following potential drawbacks, which in turn are advantages for introducing the absorbent into the scrubber circuit.

1. *Interference with boiler operation.* One of the major fears in regard to the dry limestone injection process is that the large amount of limestone injected will interfere with boiler operation, by excessive slagging on boiler surfaces, interference with fuel combustion or heat transfer, increased erosion, and perhaps other ways not yet considered. The wet process has the advantage that much less limestone will be injected; however, even the stoichiometric amount constitutes a very large introduction of foreign material into the boiler, roughly equivalent to doubling the solids content of the gas stream. Moreover, the reduced amount of limestone added as compared to the dry process might actually have an adverse effect, by reducing the slag fusion temperatures and thereby aggravating the slagging problem.

The possible adverse effects of limestone injection are discussed in detail in the dry process report. Experience from large-scale tests will be required to clarify the situation. In the current plant tests of limestone injection-scrubbing, no particular problems with boiler operation have been encountered. However, a much longer period of operation will be necessary for any conclusions. The TVA-NAPCA dry process test program will include a 6-month continuous run to evaluate the problem.

2. *Increased scaling in scrubber.* Calcium sulfate and sulfite are notorious for their propensity to precipitate from solution as an adherent scale on process equipment. The general scaling problem in limestone–wet scrubbing is discussed in detail in a later section. For the present discussion on effect of absorbent introduction point, it will suffice to say that boiler injection necessarily brings the absorbent directly into the scrubber with the gas, whereas if limestone is introduced into the scrubber circuit it can be added after the scrubber. This could have an important effect on scaling because as noted earlier, the pH increases at the point where the lime or limestone first comes in contact with the scrubber slurry, and as a result calcium sulfite and sulfate are precipitated. If this occurs in the scrubber, scaling can well result. If it occurs in the slurry circuit after the scrubber, it can be made to take place in a hold tank on the surfaces of existing crystals.

As will be discussed later, the method developed in the Howden-ICI process for preventing scaling was quite effective, whereas the Union Electric scrubber has scaled severely. The difference may be in the point of lime addition, for the British introduced it after the scrubber and before a hold tank.

Hence there is the possibility that the injection-scrubbing process may have a considerable and unavoidable disadvantage because of calcium salt deposition in the scrubber. Further plant experience is necessary to evaluate the problem.

3. *Heat required for calcination.* Although injection into the boiler is an economical way to calcine limestone, mainly because equipment requirement is low, the amount of heat required is the same as if the calcination were done in separate equipment (except for possible heat loss in off-gases in separate calcination). In contrast, if limestone is introduced directly into the scrubber circuit, the heat requirement is eliminated. The advantage for scrubber addition is not large but is significant; from figures given in the dry process report, the saving is estimated at $0.036 per ton of coal (for 3.5% sulfur in the coal and $0.26 per million B.t.u.).

4. *Loss of unreacted absorbent in solids disposal.* In injection-scrubbing, the sidestream to the disposal pond is withdrawn after the absorbent is introduced into the scrubber circuit rather than just before the introduction as in the Howden-ICI process. Therefore there may be considerably more loss of unreacted absorbent to waste. The loss would depend on rate of lime dissolution, rate of slurry recirculation, and other factors.

It is concluded that there are not enough definitive data available to favor one process to the exclusion of the other. Therefore, conceptual designs for both limestone injection-scrubbing and limestone–scrubber addition will be presented in this report.

Point of absorbent introduction is also important within each process. For example, in the injection-scrubbing method, the limestone can be added with the coal or injected through the boiler wall above the flame area; addition with the coal (in pulverized-fuel boilers) is more convenient and probably more economical in most cases. For the dry injection method, where operation must be carefully optimized for maximum sorption in the boiler, addition with the coal is questionable because of the danger of "dead-burning," that is, loss of porosity because of the high temperature to which the limestone is exposed during calcination.

Dead-burning probably is less of a problem in the wet process, because rate of dissolution in the scrubber liquor is the main consideration rather that rate of gas permeation into a porous particle. However, it may well be that dead-burning, which involves growth of CaO crystallites to a size larger than in normal calcination, could slow down dissolution rate to such a degree that absorbent utilization would be reduced. Further data are needed. As noted earlier, the limestone will be injected with the coal in one of the Kansas Power and Light installations.

Choice of point of addition when the absorbent is added to the recirculating slurry presents less of a problem. The introduction point should obviously be after the slurry has left the scrubber in order to reduce calcium salt precipitation in the scrubber.

Dust Removal

The ability of a wet scrubber to remove dust as well as sulfur oxides gives limestone–wet scrubbing a considerable advantage over the dry system for sulfur oxide removal. Eliminating the usual dust-removal equipment gives a major reduction in power plant cost that can be credited to the wet-scrubbing system.

When a new plant is involved, the cost reduction is a very strong incentive to combining removal of dust and sulfur oxides, even though the presence of the dust complicates design and operation of the scrubber somewhat. The main disadvantage is that without an electrostatic precipitator, the scrubber would have to be operated continuously for dust removal and the option for intermittent sulfur dioxide control would be complicated, as discussed below.

When an existing plant is in question, most of the ash and lime could be removed in the existing dust-collection equipment, leaving the scrubber only that part that passes through the dust collectors. In this case the lime-ash mixture caught in the dust collectors would have to be transferred to the scrubber, which probably could be easily arranged since the mixture could be dumped directly from the dust collector into the scrubber recirculating system. The question is whether continued operation of the dry dust collectors would be economical as compared with removing them and putting the full dust load on the scrubber.

Wet scrubbing for dust removal is seldom used in power plants, mainly because of high pressure drop and gas buoyancy loss. In the present case, however, these drawbacks must be accepted anyway in removing the sulfur dioxide. The remaining disadvantages of eliminating the dry collectors are:

1. High dust loading on the scrubbers, which makes some scrubber types unsuitable and may introduce operating difficulties.
2. High solids content in the recirculating slurry. This assumes that if the precipitator were retained the ash-lime mixture would be fed into a delay or settling tank and a thin slurry of controlled solids content drawn off for feed to the scrubber. Control of solids content was found important in the Howden-ICI work.
3. Possible interference with intermittent operation. If the existing dust collectors were left in place, the scrubbers could be shut down since the existing equipment could take over the dust-removal function.[4] Otherwise continuous scrubber operation would be necessary to remove dust, the same situation as when dust and sulfur oxide removal are combined in a new plant. Costs could still be reduced by feeding lime only when weather conditions were poor for plume dispersion, but during the nonfeeding periods the scrubbing water would rapidly become acidic, in which case resistant construction material would be required and it would probably be necessary to recycle the acidic scrubber effluent from the pond.
4. Scaling probably would be aggravated. If the dust collectors were left in, the lime would not go directly into the scrubber and therefore the calcium salt precipitation in the scrubber mentioned earlier would be avoided.

If limestone is introduced directly into the scrubber rather than injected into the boiler, there is even less justification for removing the dust collectors. In this system the scrubber does not have to be designed for dust removal (although residual dust from the dry collectors must be considered in scrubber operation) and the ash is kept separate from the lime (which may improve scrubber operation and possibly make the reaction product more useful as a byproduct).

The main advantage of removing the dry collectors is eliminating the cost of operating them–assuming that the scrubber operating cost is not increased to an equal or higher degree by the dust loading. These factors are analyzed in the cost section of this report.

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[4] Assuming that a practical equipment arrangement could be worked out; see later discussion.

Scaling

Whenever solutions containing calcium sulfate or sulfite must be handled in a process, trouble with scaling can be expected. It is a major problem in the fertilizer industry where, in the production of phosphoric acid by the wet process, calcium sulfate formed by reaction of calcium phospahte ore with sulfuric acid crystallizes in lines and on equipment surfaces to the extent that the plant must be shut down periodically for "boiling out." Scaling was the only major problem encountered in development of the Howden-ICI process and the current experience at Union Electric indicates that it will be also in process developments in this country.

Although solids in the recirculating scrubber suspension serve as nuclei for crystal growth, nucleation can also occur on scrubber surfaces with resulting scaling, particularly since both calcium sulfate and sulfite form supersaturated solutions from which precipitation is slow even when the degree of supersaturation is high. In other similar systems the concentration of dissolved salts in the recycling liquor can be reduced by cooling outside the scrubber and supersaturation within the scrubber thereby avoided. However, the solubilities of both calcium sulfate and calcium sulfite have an inverse temperature coefficient and the method therefore is not applicable.

The composition of several samples of scale has been reported by Lessing (13).

$Ca(OH)_2$	Nil
$CaCO_3$	0-19.7%
$CaSO_3$	0.8-14.3
$CaSO_4$	62.8-97.2

A technique was developed in the Howden-ICI process (13) that was quite effective in control of scaling. The main feature was provision of a delay tank after the scrubber in which the slurry was held for a period long enough for the supersaturation to dissipate on the surface of suspended crystals; the effect of retention time and content of suspended crystals is shown in Figure 2. It was also necessary to (1) maintain a minimum content of suspended crystals (3% each of calcium sulfate and sulfite) in the

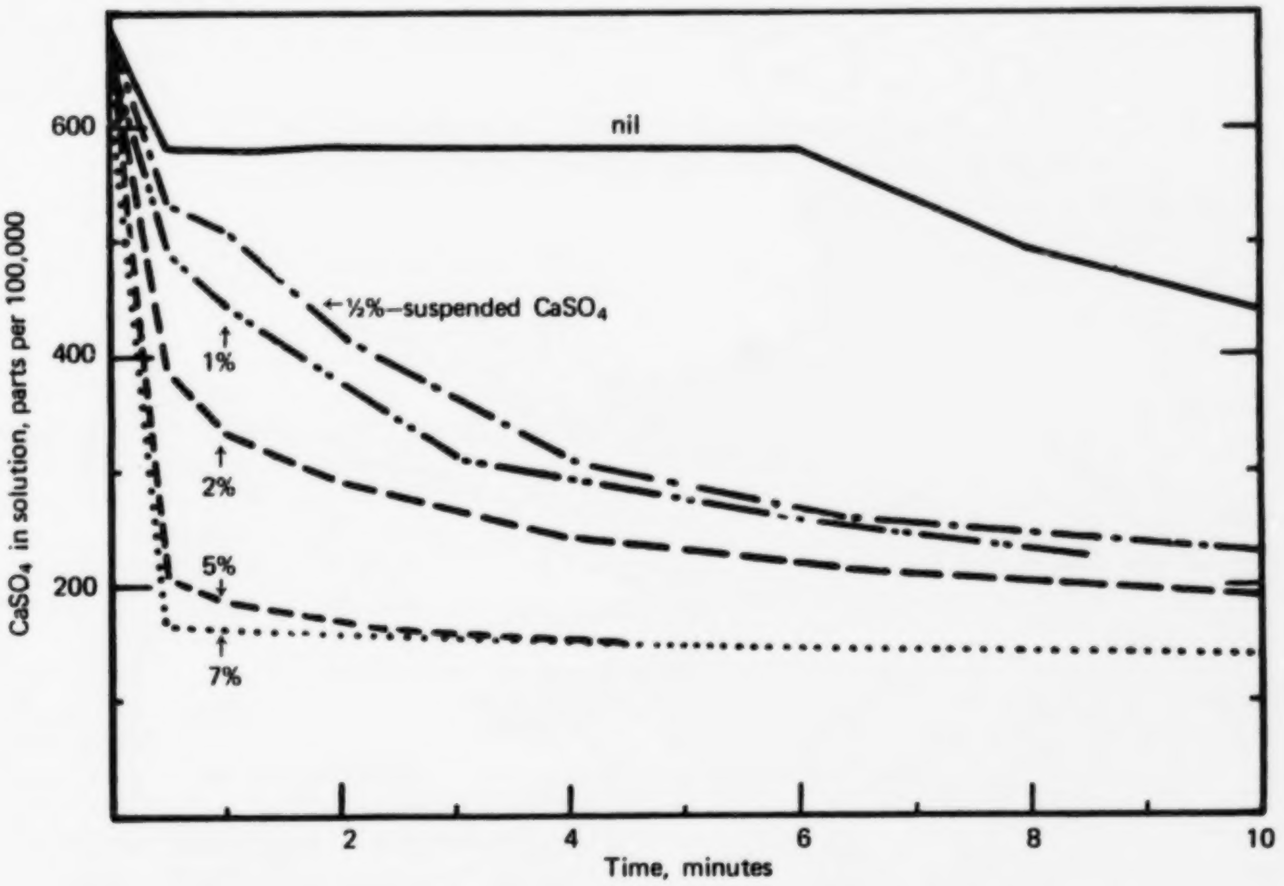

Figure 2. Desupersaturation of calcium sulfate solution at 50° C. by suspended $CaSO_4 \cdot 2\ H_2O$

slurry returning to the scrubber as a means of providing surface for salt deposition at all points in the system; (2) use a relatively high slurry recirculation rate to minimize buildup of supersaturation in the scrubber; and (3) provide a special "scale resistant" packing (suspended wood plates) in the lower section of the scrubber to prevent trouble from the small degree of supersaturation developed at that point.

By use of this technique, the Howden-ICI workers were able to eliminate completely the severe scaling that had been their main problem in developing the process. The plant at Fulham was operated for an extended period with no evidence of scaling. It should be noted, however, as pointed out earlier, that much of the freedom from scaling may have been due to the fact that the absorbent was introduced between the scrubber and the delay tank rather than directly into the scrubber. Absorption of sulfur oxides in the scrubber builds up the calcium sulfate and sulfite concentration but the accompanying decrease in pH increases solubility of sulfite and helps avoid supersaturation in the scrubber. Introduction of the lime or limestone increases pH, decreases solubility, and induces precipitation or supersaturation. If the introduction is just before the delay tank, precipitation and dissipation of supersaturation occur at a point where they cause no trouble.

The scaling reported (11) in the current tests at the Union Electric installation possibly is caused by introduction of the lime directly into the scrubber. Or other factors identified by the British workers may be involved; further data from the tests are needed to clarify the situation. The mobile-bed scrubber type in use at Union Electric was tested in the pilot plant programs at Detroit Edison and Wisconsin Electric. Scaling was not reported to be a problem; apparently the movement of the marbles or spheres against each other prevents scaling in the bed, but good liquid distribution is necessary to prevent inactive bed areas in which solids deposition could occur. Moreover, scaling can occur in areas not in contact with the mobile bed.

Corrosion

Corrosion can be a major problem in wet scrubbing to remove sulfur oxides. Unless the pH of the scrubber solution is carefully controlled at nearly neutral levels, attack of exposed metal surfaces is serious unless expensive alloys are used. Corrosion of duct work and scrubbers was said to be a major problem in the early English work; however, no corrosion data are given in the various publications. The general statement was made that mild steel is satisfactory for the scrubber shell if the pH is controlled at a minimum of about 6.2 at the scrubber outlet. A more resistant material was required for scrubber packing, valves, pumps, and lines (at some points).

Corrosion was not a problem in the pilot plant work at Wisconsin Electric; the scrubber was of fiber glass-reinforced polyester construction, a material resistant to both acid and alkaline attack. Use of this material for the large scrubbers required in commercial application of the process would be complicated by structural problems. However, one scrubber manufacturer indicated ability to produce such units. Costs would be somewhat higher than for coated mild steel.

Corrosion data were not reported for the CE-Detroit Edison tests. The stack gas treatment facilities installed by Union Electric and Kansas Power and Light include scrubbers fabricated from mild steel and coated internally with a bitumastic lining. The support trays, internal piping, and spray nozzles are of stainless steel construction. Equipment external to the scrubber, including recirculating piping, pumps, settling tank, and line to the pond, are of mild steel.

Limestone Effectiveness

In the dry injection process, the properties of the particular limestone used are quite important because of the considerable differences that have been found between limestone types in effectiveness for sulfur oxide sorption. The reasons for this are not fully known; impurities apparently play a part, possibly by blocking absorptive surfaces, and some limestones appear less resistant to dead-burning than others. In any event, selection and procurement of an effective limestone is a major problem in the dry process.

It has been generally considered that limestone properties would not be nearly so important in the wet-scrubbing process, particularly in the injection-scrubbing version. Although some sorption occurs in the boiler, it is incidental and actually is not needed; the boiler needs only to accomplish calcination, for the scrubber is quite capable of doing all the sulfur oxide removal, as in the British work.

It is conceivable, however, that the same properties which cause poor sorption in the boiler could also reduce the dissolution rate of the calcined limestone when it reaches the scrubber. In the boiler the problem is developing a large sorption surface in the lime particles and getting the gaseous sulfur oxide into the particle interior and into contact with the reactive surface; any limestone property that reduces the surface or interferes with mass transfer is undesirable. To some extent the same may be true in the scrubber, where the problem is to get the scrubber solution into contact with reactive surfaces of the lime to get it dissolved. Insoluble impurities fused onto the CaO surface, low porosity, large crystallite size of CaO, and a shell of calcium sulfate on the particle all could delay dissolution.

Offsetting this is the relatively long retention time in the scrubber loop; a lime particle should have ample opportunity to dissolve because it circulates in the system until it happens to be caught in the sidestream going to the solids-disposal pond. As noted earlier, however, the fact that in the injection-scrubbing method the sidestream is withdrawn shortly after introduction of the lime into the system increases loss of unreacted absorbent to waste and therefore would aggravate the adverse effect of slow dissolution.

No data are available on the question. The only integrated test of injection-scrubbing was in the work of Combustion Engineering and no tests of limestone type were reported. Further test work to evaluate differences in limestone effectiveness are needed.

More information is available on effectiveness of various limestones for the system involving absorbent addition to the recirculating slurry. During the Howden-ICI work, it was found that precipitated calcium carbonate was a satisfactory absorbent but some natural chalks, even when ground to minus 200 mesh, were not sufficiently reactive unless used in considerable excess. In contrast, finely ground limestone was a good absorbent in early pilot plant tests at TVA, but precipitated calcium carbonate gave poor results. In recent bench-scale tests at TVA, a chalk-type limestone absorbed more sulfur dioxide (by a factor of about 3), under the same scrubbing condition, than a hard, dense, limestone of the prevalent type. In the pilot plant work at Wisconsin Electric, pulverized limestone was an effective absorbent.

These results indicate that differences in limestone properties will be important in the scrubber circuit addition method. It may be possible to use any type of limestone if provisions are made to get it into solution before the slurry enters the scrubber, but the conditions needed for accomplishing this may vary between limestones. More research is needed.

Solids Disposal and Water Pollution

The make of calcium sulfate and sulfite must be removed from the scrubber circuit and conveyed to the disposal point. In so doing, problems occur both in transporting the waste and in disposing of the accompanying water.

In the Howden-ICI process, the solids concentration in the circulating system was controlled in the range of 10 to 15% by diverting a purge stream (from a point between the scrubber and the point of lime addition) to a conical settling tank from which an underflow containing 33 to 40% solids was withdrawn. The underflow was then filtered on a rotary drum to reduce the moisture content of the solids to 30-45%, depending on particle size. Partially clarified overflow from the settling tank and filtrate from the filter were returned to the process; solids were discarded to a waste pile. Partial clarification was considered desirable because the smaller particles returning to the scrubber served as seeds for crystal growth. Filtration rates were about 40 pounds per hour per square foot.

Sludge disposal was not studied during Combustion Engineering's experimental work but concentration studies were made by equipment manufacturers with mixtures of fly ash and reaction products from the test program. The Meramec installation includes a thickener for concentrating the slurry to about 35% solids before it is discharged to the fly ash pond; thickener overflow is returned to the scrubber. At the Kansas Power and Light Lawrence station the entire scrubber effluent is pumped to a settling pond and the overflow returned to the scrubber.

No information is available on such points as minimum solids content for the slurry to avoid flow troubles in the sluice line, erosion of pumps and lines, and piling characteristics of the dewatered sludge. These are all important problems in the disposal of gypsum from phosphoric acid plants in the fertilizer industry, where the general practice is to sluice to a pond and by use of dikes to build the gypsum pile to heights as much as 100 feet above ground level.

The solids could be dewatered at the scrubber and conveyed to the disposal pile by solids-handling equipment, a procedure that would eliminate pumping problems. The method is quite expensive, however, and is seldom practiced in disposing of gypsum from phosphoric acid plants.

A major problem in solids disposal is what to do with the sluice water decanted or drained from the pond. It can either be pumped back to the power plant and reused in the scrubber or allowed to flow into an adjacent watercourse. Recycling is much more desirable from the standpoint of water pollution because it eliminates any release of salts to watercourses other than from seepage or accidental pond overflow. From the process standpoint, however, recycle is undesirable both because it is more expensive and because recycling the saturated solution may introduce operating problems.

Hence water pollution must be weighed against the cost and adverse process effects of recycling the water.[5] The cost of recycling can be estimated; it depends on the distance from scrubber to pond. The effect on operation, however, is difficult to predict. Increased scrubber scaling might well result because the amount of fresh water added to the recirculating slurry would be reduced; the fresh water makeup normally used to replace the pond water drained to the watercourse would be eliminated. Hence the slurry entering the scrubber would have a higher degree of supersaturation and scaling would be more difficult to prevent.

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[5] It should be noted, however, that recycling is not a perfect solution to the pollution problem. In the phosphoric acid industry, pond water recycle is often practiced but there is still trouble with pond dike breakage and with pond flooding by heavy rains, with resulting escape of the pond water to the watercourse.

It should be noted in this respect that in the pilot plant work reported by Pearson (4) most of the water was removed from the solids and recycled. The resulting "mud" contained only 35% water as compared with 60% in the heaviest slurry considered suitable for pumping in the present study. The overall fresh water makeup was 0.62 ton per ton of coal burned, of which about 0.1 ton was from water added to replace that lost in the solids; the remainder replaced that consumed in humidifying the stack gas. The pilot plant gave no scaling trouble under these conditions, even with only 0.1 ton of water leaving the system with the solids and replaced with fresh water; thus elimination of this small amount by recycling might not have any significant adverse effect.

Scaling in the sluice line might also be aggravated by continual use of saturated solution as sluice water. If the water is not returned, excess fresh water could be used in sluicing to prevent scale; however, it should also be possible to sluice with the fresh water makeup in a recycle pond water system.

It is assumed that recycling the pond water is the preferred course and that such a "closed loop" system will be used unless the situation is such that the cost and operating difficulty, coupled with a relatively low and acceptable outflow of polluting salts, makes recycling unjustified. The problems in deciding whether or not recycling is justified are reviewed in Appendix B.

Stack Gas Reheat

One of the major problems to be dealt with in the wet-scrubbing process is loss of stack gas buoyancy as a result of cooling.[6] Much emphasis has been placed in recent times on use of high stacks (up to 1200 ft.) to improve dispersion and reduce the possibility of localized high pollutant concentrations. A reduction in stack gas temperature by wet scrubbing will reduce both the momentum and buoyancy which cause the plume to rise above the stack after it is emitted. Thus, the effective stack height and the plume dispersion will be reduced by wet scrubbing. This should not be serious if essentially all of the objectionable components are removed in the scrubber but such efficiency is not likely, and, in any event, upset in operating conditions could release large quantities of pollutants, in which case the safeguard of good dispersion through high discharge would be desirable.

Humidification of the stack gas is also objectionable because condensation may occur and cause formation of a visible plume which could have an undesirable psychological impact on the public. The importance of the visible moisture plume is difficult to assess. Power plants have been operated with such plumes (in the U. S. and France), presumably without difficulty. Also, large cooling towers, which are becoming rather common, produce low-level plumes. It is concluded that the plume is undesirable but that, other things being equal, no considerable expenditure in eliminating it is justified.

These problems have been generally considered so serious that after the first studies in England and the U. S. the wet-scrubbing method fell into disfavor and emphasis was shifted to dry methods. It is only recently that the work of Combustion Engineering, Wisconsin Electric, and others (including work in France and Czechoslovakia on scrubbing with aqueous ammonia solutions) has brought the method back into consideration.

The obvious solution to the problem is to reheat the gas before release to the atmosphere, but this apparently has been generally regarded as too expensive. No cost figures have been published, however, to show that the cost would be intolerable. Instead, cost estimates published recently by Combustion Engineering indicate that both capital and operating cost would be acceptable (8).

Very little test work on reheating has been reported. Combustion Engineering operated the pilot plant with a finned-tube exchanger inserted in the scrubber off-gas duct. Solids accumulated on the surface but they were easily removed by water washing. The system at Union Electric includes finned-tube heat exchangers to extract heat from the stack gas ahead of the scrubber (after the air heater) and to reheat the exhaust. A circulating ethylene glycol-water mixture is used to transfer the heat. Design conditions provide for cooling the inlet gas to the scrubber from 282 to 232°F. and for reheating from 123 to 173°F. At the Kansas Power and Light installation, the scrubber exhaust will be reheated by heat exchange with boiler feed water.

There is general agreement that the cooled gas should be reheated but no agreement at all on the degree of reheating necessary or desirable. The consensus seems to be that it need not be reheated to the original temperature but that it should be heated at least to a temperature high enough to avoid a visible plume under most atmospheric conditions.

The reheating problem is a complicated matter. The gas is not only cooled in the scrubber but the composition is changed radically; most of the sulfur oxides are removed (in an efficient system), part of the nitrogen oxides are removed,[7] and the water vapor content is increased to

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[6] Normally to the wet bulb temperature, about 125° F.

[7] Data on nitrogen oxide removal reported by various organizations are as follows:

Organization	Removal of NO_X, %
Howden-ICI	63-65
UOP-Wisconsin Electric	19-23
CE-Detroit Edison	20 (approx.)

The reason for the large difference between results is not known. The burden of proof seems to be on the Howden-ICI data as they are much higher than reported for other aqueous scrubbing methods.

saturation. The problem is how to predict maximum ground-level pollutant concentration in the power plant vicinity for the changed conditions in regard to gas composition and temperature.

Effect of Stack Gas Reheat Temperature on Plume Dispersion: TVA has carried out comprehensive studies of plume dispersion from coal-fired, electric generating plants. Based on this experience, the effect of stack gas temperature on maximum sulfur dioxide concentration at ground level was determined for a 200-megawatt unit with a 300-foot stack. From full-scale field studies (16), the plume rise above stack top was found to be:

$$\Delta h = C\left(\frac{F}{\bar{u}^{3}}\right)^{.27} \qquad (8)$$

where

Δh = plume rise above stack top
C = stability coefficient
F = flux due to buoyancy and momentum = $g V_s r^2 \left(\frac{\Delta T}{T_a}\right)$
g = acceleration due to gravity
V_s = stack gas exit velocity
r = stack exit radius
ΔT = excess stack gas temperature over ambient temperature
T_a = ambient air temperature
$\bar{u}$ = average wind speed

The groundline SO_2 concentration at any point downwind is given by the equation:

$$\chi = \frac{Q}{2\pi C_y C_z \bar{u} x^m} \exp\left[-\frac{1}{2x^m}\left(\frac{y^2}{C_y^2} + \frac{z^2}{C_z^2}\right)\right] \qquad (9)$$

where

χ = SO_2 concentration
Q = SO_2 emission
C_y, C_z, and m = diffusion coefficients
$\bar{u}$ = average wind speed
x = distance downwind
y = distance crosswind
z = height of plume (above the ground, center of cross section)

By means of these two equations, the maximum expected groundline sulfur dioxide concentrations were determined for (1) stack gas temperatures (at scrubber or reheater outlet) ranging from 125 to 250° F. at 25 intervals, (2) scrubber efficiencies of 95, 90, 85, and 80%, (3) the plant operational conditions shown in Table 3, and (4) wind speeds ranging from 6 to 30 miles per hour. The critical wind speed (speed at which maximum SO_2 groundline concentration occurs) ranged from 8 to 26 miles per hour for the various combinations of conditions. The calculations were based on neutral to slightly unstable atmospheric temperatures, that is, near normal decrease in temperature with altitude increase. It was also assumed that the gas contained no mist at the point of emission from the stack, and that therefore evaporative cooling of the plume would not adversely affect dispersion.

Results of the calculations are shown in Table 4; the values shown are based on a wind speed of 18 miles per hour, as most of the conditions gave critical wind speeds close to this value. The wind speed was held constant to illustrate the effect of stack temperature alone on plume rise.

The main conclusion from Table 4 is that maximum ground concentration is increased by about 85% if the gas is emitted at scrubber exit temperature (without reheating) rather than at normal stack temperature. At high degrees of removal, however, this increase is not highly significant because the concentration is still quite low even after the 85% increase. For example, at 95% removal and 125° F. stack temperature the maximum ground concentration is 0.015 parts per million, which is equivalent to 91% reduction in ground concentration as compared with the

Table 3. Plant Operating Data: Colbert Steam Plant (Unit 1)

Rated capacity, mw		200
Stack height, ft		300
Coal burned, lb/hr		150,000
Stack diameter, ft		16.5
Ambient temperature, °F. (assumed)		60
Stack gas temperature, °F.		
Entering stack	Leaving[a] stack	Stack gas velocity, fps
310	296.5	42.26
250	236.5	38.91
225	211.5	37.51
200	186.5	36.12
175	161.5	34.72
150	136.5	33.32
125	111.5	31.92
Coal analysis		%
Moistura		7.9
Volatile matter		34.2
Fixed carbon		42.4
Ash		12.0
Sulfur		3.5
		100.0

[a] Temperature checks show approximately 13.5° F. loss between inlet and outlet of stack.

0.16 parts per million for normal power plant operation without scrubbing. At lower degrees of removal, however, the 85% increase becomes more significant. Since the ground concentration is directly proportional to the amount of sulfur dioxide left in the gas (see equation 2), it follows that at 50% removal (10 times as much SO_2 remaining in the gas as for 95% removal), the 0.015 parts per million concentration becomes 0.15 parts per million, or practically the same as the 0.16 parts per million at normal stack temperature but with no sulfur dioxide removal. In other words, 50% removal gives little or no benefit if the gas is not reheated.

Reductions in ground concentration at various degree of reheat are shown in Figure 3. This covers low degrees of removal and thus shows the effect of gas cooling on ground concentration of gas constituents such as nitrogen oxides that are removed to only a limited degree in the scrubbing process. If 30% removal of nitrogen oxides is assumed, the gas would have to be reheated to 190° F. to avoid an increase in ground-level concentration as compared with normal power plant operation.

The graph also allows evaluation of reheating by use of a scrubber bypass (in systems where an existing electrostatic precipitator is left in operation). If we assume bypass of 25% of the gas (at 310° F., the air heater outlet temperature) and 90% removal of sulfur dioxide from the remainder, the resulting gas temperature would be about 170° F. and the net degree of sulfur dioxide removal 67.5%. The resulting reduction in ground-level concentration would be only about 50%. In contrast, scrubbing all the gas but with no reheat would give a concentration reduction of 82%. Hence bypassing of 310° F. gas does not appear to be an acceptable reheating method. Hotter gas from ahead of the preheater could be used but taking heat from this source would incur a fuel cost and would be equivalent to direct heating of the gas (see later cost estimate on this); moreover, ducting would be a problem and the thermal balance in existing plants would be affected.

Table 4. Maximum Groundline Concentrations of Sulfur Dioxide[a] at 18-mph Wind Speed

Stack gas temp.,[b] °F.	Plume rise above top of stack, ft	SO_2 concentration, ppm, at indicated % of SO_2 removed				
		95	90	85	80	0[c]
310	402	0.008	0.016	0.024	0.032	0.16
250[c]	363	0.010	0.020	0.030	0.040	0.20
225	345	0.011	0.021	0.032	0.042	0.21
200	341	0.011	0.022	0.033	0.044	0.22
175	318	0.012	0.024	0.036	0.048	0.24
150	292	0.013	0.026	0.039	0.052	0.26
125[d]	259	0.015	0.029	0.044	0.058	0.29

[a] Based on plant data from table 3. In addition: total SO_2 in gas, 126 tons/day; SO_2 in gas, about 0.3%; approximately 90% of the sulfur in coal leaving the system in the stack gas.

[b] At exit of last operating unit before stack; further loss of 13.5° F. in ducts and stack.

[c] Highest practical reheat temperature.

[d] Without reheat.

Calculations were made for other stack heights also. For normal power plant operation, cutting the height from 300 to 150 feet increased maximum ground concentration by about 75%. At 80% removal and without reheat (stack temperature of 125° F.), the increase was 100%. Conversely, if the stack height were 600 feet instead of 300, the influence of gas temperature would diminish; ground-level concentration would be increased only about 45% by releasing the gas at 125 rather than 310° F., as compared with 85% increase with the 300-foot stack. At lower wind speeds, which are generally more critical in regard to ground fumigation from high stacks, release of gas at 125° F. would be even less significant; the increase in ground concentration over release at 310° F. would be only about 25%. Thus as stack height is increased above the 300 feet assumed, reheat becomes less important.

It is concluded that the gas should be reheated after scrubbing but that the degree of reheat desirable may vary widely depending on the particular situation. As noted earlier, it is assumed in this study that reheating beyond 250° F. would not be attempted because of the relatively high cost for going to higher levels. The minimum acceptable temperature probably is that required to avoid presence of mist in the gas as it leaves the stack (see next section), which would vary with power plant duct and stack design and with efficiency of the mist eliminator. Between these two levels the appropriate degree of reheat probably will vary depending on local factors and on degree of sulfur dioxide removal for which the unit is designed. Relatively low sulfur dioxide removal should be coupled with relatively high degree of reheat to keep ground concentration at a reasonably low level; for example, 70% removal would require reheat to about 215° F. to avoid ground concentration reduction from falling below 60%.

Effect of Evaporative Cooling: In the scrubber, hot unsaturated gas is intimately contacted with the scrubbing medium to promote rapid sulfur dioxide absorption; as a result water is evaporated from the liquid and the gas becomes saturated. After saturation, further loss of heat may occur by contact with cooler surfaces, in which case water vapor condenses in the form of mist droplets. The amount of condensate formed is a function of the heat loss from the gas after saturation. In addition, some droplets will be present because the mist eliminator does not remove all of the mist formed in the scrubber. Also, condensation will occur in the stack when saturated gas is cooled by heat loss from the stack.

After the gas leaves the stack, further heat loss to the ambient air takes place and more condensation occurs. Then as the plume rises higher and, in many cases, is

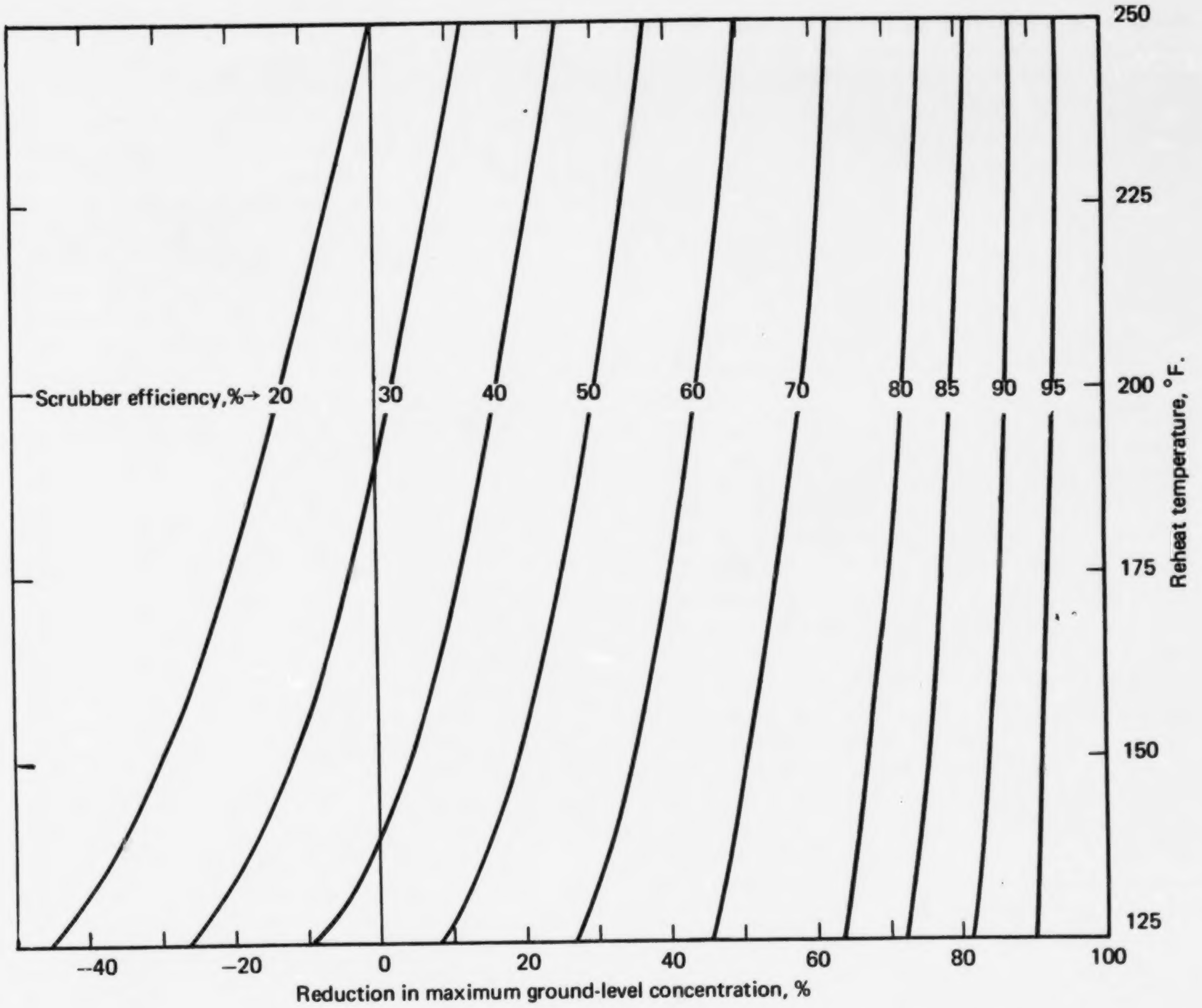


Figure 3. Effect of reheat temperature on ground-level SO_2 concentration at various scrubber efficiencies (Base: Maximum ground-level concentration at 310° stack temperature and without scrubber)

subjected to wind energy, mixing with ambient, unsaturated air occurs and the dilution soon gives an unsaturated mixture, still within the confines of the plume. The resulting evaporation, which continues until all the droplets are evaporated and the plume becomes clear, cools the plume—possibly to a temperature below that of the surrounding air. If this develops the plume will have a "negative buoyancy" and may sink to the ground faster than if the evaporative cooling had not taken place.

The situation is different from that in normal power plant operation, where the stack gas is quite unsaturated at the stack exit temperature. The dew point is so far below the stack temperature that dilution by unsaturated ambient air drops the dew point further, to below ambient temperature, before condensation can occur.

Scorer (17, 18) has treated the effects of heat of condensation in offsetting the loss of buoyancy resulting from evaporative cooling. For condensation occurring after the gas leaves the stack, it is obvious that the amount of heat evolved during condensation is equal to that absorbed during evaporation. Hence the effects cancel each other and the final temperature should be the same as for the usual situation—an unsaturated gas leaving the stack. However, the heat of condensation makes the gas hotter than it otherwise would be and therefore it rises higher than would an unsaturated gas. Thus any downward movement resulting from evaporative cooling would begin at a point higher than unsaturated gas would reach. The faster rate of cooling presumably would bring the plume down faster so it is difficult to say what the net effect would be. It seems possible, however, as pointed out by Scorer, that a saturated plume would actually disperse better because of latent heat effects.

The situation in wet scrubbing is complicated by the fact that some of the condensation occurs before the gas leaves the stack. Since the resulting mist does not contribute any latent heat to the gas after it leaves the stack, its heat of evaporation all goes to cooling of the plume. As a result the ultimate plume temperature should be lower than for an unsaturated plume leaving the stack at the same temperature. Moreover, the mist increases bulk density and hence reduces buoyancy.

Scorer treats this problem in terms of stack exit temperature as compared with wet-bulb temperature, which is a measure of the degree of condensation because any difference between these temperatures is due to heat loss through duct and stack walls and this cooling is the cause of the condensation. Scorer developed a nomogram (Fig. 4) for quantitative estimation of the cooling of a wet plume by evaporation of water condensed before emergence—at various emission temperatures. Plotted isopleths show the degree of cooling by evaporation and can also be used to determine the plume temperature after dilution. For example, with 100-fold dilution and a gas wet-bulb temperature of 125°F., a 10°F. temperature drop in the saturated gas before emission would result in about a 1.3°F. decrease in plume temperature; with less dilution, say 10-fold, the plume would be cooled 13°F. Based on results of plume dispersion studies, the 100-fold dilution is probably representative of typical conditions and the effect of negative buoyancy would be insignificant.

However, mist escaping through the mist eliminator would add to the problem, particularly since it does not contribute any heat of condensation at all. Mist from the scrubber may well have been the cause for the situation in the tests at the Bankside station in London, where, according to Scorer, the plume could "almost invariably be seen to obscure the chimney tops and usually reached the ground within a few hundred yards when ambient turbulence was not large."

Scorer's study is one of very few on the subject. Inquiries to leading meteorologists in the air pollution field indicate that the effect of evaporative cooling on plume rise has received only minimal attention to date. However, with the advent of large, natural-draft cooling towers, it is likely that the subject will receive increasing attention.

It is concluded that evaporative cooling may have an adverse effect on dispersion if there is any mist in the gas when it leaves the stack. Present information, however, does not allow a quantitative treatment of the resulting plume cooling. Any evaporative cooling is likely to make the formulas in the preceding section inaccurate, in the direction of predicting lower ground-level concentrations than would actually occur.

A further conclusion is that the gas should be reheated at least enough to ensure absence of mist at the point of emission from the stack. Heating to a higher temperature would not help in regard to the evaporative cooling effect, but would, of course, increase plume buoyancy as discussed in the previous section.

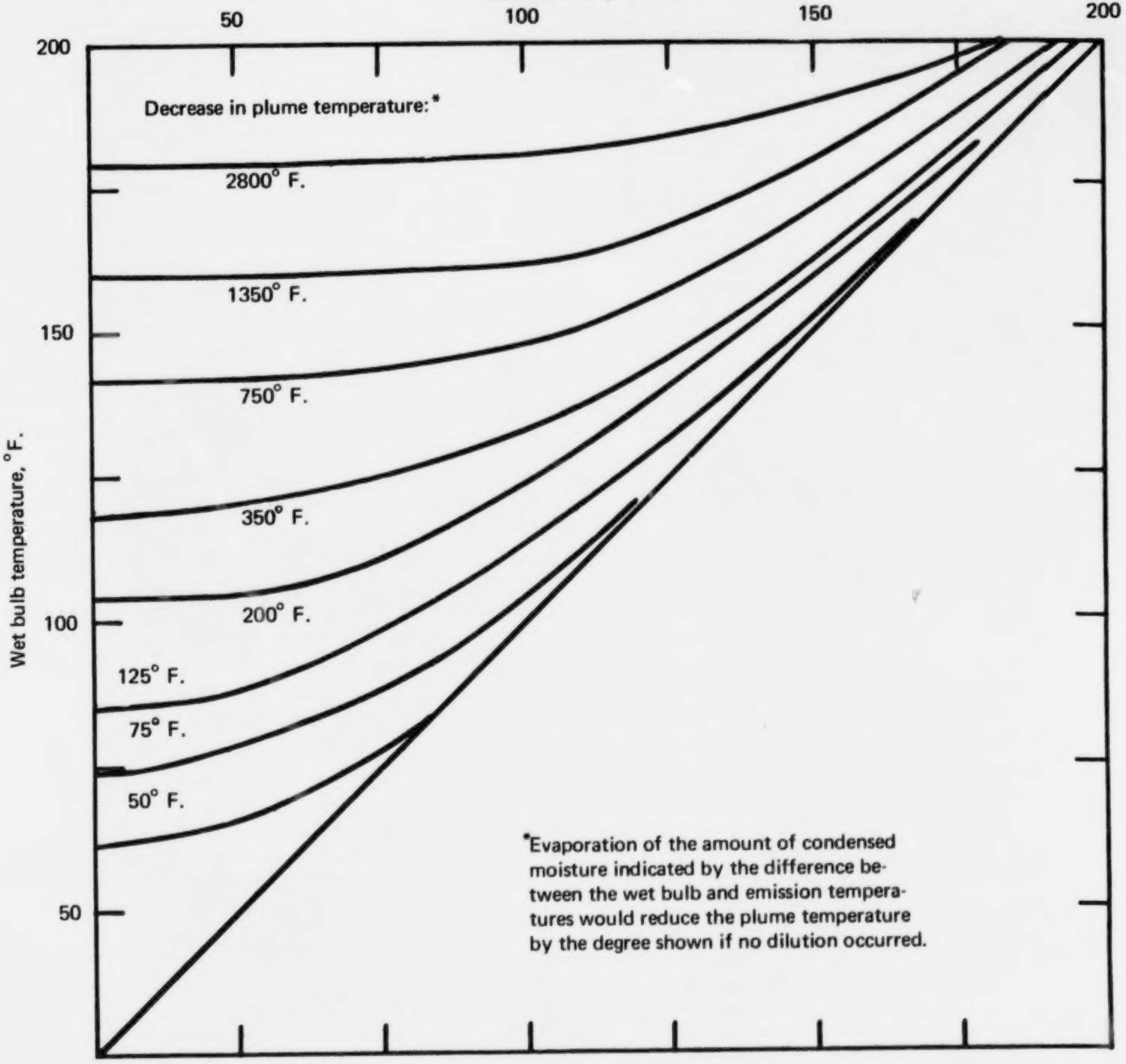
Intermittent Operation

In the dry process study it was assumed that limestone injection could be used by power plants in two ways—continuously where local conditions dictate and occasionally where the main need is to avoid high ground concentrations that develop only when atmospheric conditions are adverse to good dispersion from the stacks. Since the latter is a very good way to reduce the cost of sulfur oxide control, amenability to intermittent operation is a highly desirable characteristic of any control process.

Operating flexibility is somewhat limited in the limestone—wet scrubbing process because the scrubber also serves as a dust collector. In new plants the scrubber would have to be operated whenever the power plant was on line because there would be no other way to remove dust. It would be possible, of course, to achieve intermittent sulfur oxide removal by the same means as in the dry process—shutting off the feed of limestone to the boiler. The scrubber would continue to remove dust with water as the scrubbing agent. As noted earlier, however, the scrubber liquor would quickly become acidic; as a result the scrubber and waste-disposal system would have to be built of corrosion-resistant material. Moreover, any mist escaping from the mist eliminator could corrode parts of the power plant after the scrubber, including the stack.

In existing plants already equipped with electrostatic precipitators, intermittent sulfur oxide removal could be accomplished by (1) connecting the precipitator in parallel with the scrubber and shifting the gas flow to the precipitator when sulfur oxide scrubbing was not needed or (2) connecting the precipitator and scrubber in series and shutting off the scrubber liquor flow to stop sulfur oxide removal.

The main drawback to parallel operation is difficulty in operating and maintaining the very large shutoff dampers required and the fact that expected leakage through the dampers would probably cause troubles in the equipment not in service at the time. Leakage of gas through the



*Evaporation of the amount of condensed moisture indicated by the difference between the wet bulb and emission temperatures would reduce the plume temperature by the degree shown if no dilution occurred.

Figure 4. Cooling of plume by evaporation of water condensed before emission (17)

damper into the precipitator could have serious consequences. The precipitator would be at or near ambient temperature when out of service and the gas leaking into it would be expected therefore to cool below the dew point of sulfuric acid. Dust leaking through would accumulate and possibly form sticky and corrosive deposits with the sulfuric acid. Also, a temperature differential might develop across the precipitator that could cause warpage of electrical collectors.

It should be noted that use of dampers is included in the conceptual design presented later, in which their function is to close off one of the two scrubbers at low power plant load (a single scrubber would be too large to operate satisfactorily at as low a load as sometimes required of power boilers). The dampers are included because there appears to be no alternative, but precipitator corrosion and caking is not involved. Gas leakage into the scrubbers may not be a major problem since they are equipped with liquid feed facilities and therefore could be flushed occasionally with water or alkaline solution during nonoperating periods.

Connecting the precipitator and scrubber in series appears to be a much more practical arrangement for intermittent sulfur removal in an existing plant. Problems with putting the precipitator in and out of service would thus be avoided and there would be no dampers to give trouble. To shut down sulfur removal, the scrubber liquor would be cut off and water introduced to flush out all salt residue. The scrubber would then dry out and ride on the line until sulfur oxide removal was required again.

Possible drawbacks to this include accumulation of residual dust from the precipitator in the packing and on surfaces. Perhaps an occasional flushing with water would prevent any trouble from such accumulation. Another potential problem would be the possibility of damage to temperature-sensitive parts of the scrubber, e.g., polypropylene sphere packing or scrubber shell coating, by the hotter gas. If this were found to be a major problem the gas could be bypassed around the scrubbers to the stack. Such an arrangement would bring back the damper problem but perhaps it could be handled as outlined above.

It is concluded that the limestone–wet scrubbing process could be operated intermittently if several anticipated problems were resolved. There is little doubt that these problems can be worked out; a program of engineering development very likely would provide suitable solutions. However, the overall plant investment would be increased in fitting the plant for intermittent operation and there would be some increase in overall operating cost that would tend to offset the saving in limestone. These extra costs could so cut into the advantage of intermittent operation that continuous operation would be preferable, with the added benefit of assured pollution control at all times.

Study Assumptions and Design Criteria

The conceptual design in this study has been developed for two basic processes:

Process A. Injection of pulverized limestone into a boiler with subsequent scrubbing of the flue gases to remove both fly ash and sulfur dioxide.

Process B. Introduction of lime or pulverized limestone directly into the scrubber recirculating slurry to remove sulfur dioxide; fly ash will also be removed in the scrubber.

The effect of dust removal prior to scrubbing through use of electrostatic precipitators is also evaluated for both processes. These four arrangements should cover most potential applications of limestone–wet scrubbing. Each process is compared under uniform conditions and in addition, the effects of variables which significantly affect costs are presented.

Plant Location

As a basis for development of cost data, the TVA Colbert power plant in northwestern Alabama (16 mi. west of Muscle Shoals) was selected. This plant is located on the Tennessee River, has an overall capacity of 1300 megawatts (one 500-mw. and four 200-mw. units), and has coal-burning boilers of the pulverized-fuel type. The same plant was used as the cost basis in the dry limestone injection study.

The Colbert location should serve adequately as a general case. The main variables in plant location are distance from limestone supply and availability of space for reacted limestone disposal. In these respects the Colbert plant is well located; limestone is available from operating quarries about 10 miles away, and disposal space is not a major problem because the plant is located in a farming area where additional space is available and probably could be purchased at reasonable cost. The same situation appears to hold for many of the power plants in the United States. The main exceptions are plants located in or near urban centers, where quarry operation may be feasible only some distance away and disposal space may command a premium or even be unavailable.

To cover plant locations not as favorably located as the one selected, the effect of freight cost on limestone delivered price was reviewed; the results are presented in the cost estimate section. Costs of disposal also were evaluated.

Plant Size

A 200-megawatt unit size was selected as the base case for the study. From the chart of power plant sizes in Figure 5, it can be seen that this is a typical midrange boiler size. Estimates were also made for 500- and 1000-megawatt boilers.

Fuel Type

Coal was selected as the fuel because it is the prevalent type. Many plants burn high-sulfur oil, however, and limestone–wet scrubbing would be an appropriate method for sulfur dioxide control. The main difference is that since oil produces very little ash, oil-burning plants have limited facilities for handling solids. Equipment for solids disposal would be new for an oil-fired plant instead of an addition to existing facilities for coal firing and therefore procurement of land for pond construction might be a problem.

Sulfur Content of Coal

The coal deposits in use vary widely in sulfur content as shown in Table 5. Three levels, 2%, 3.5%, and 5%, were selected for the study. It seems likely that coals of relatively low sulfur content will be consumed rapidly in the next few years and that the higher levels will become prevalent. The 5% level was selected as being typical of coals which might be used in processes designed for recovery of sulfur in marketable form and was included as a basis of comparison for future evaluation of recovery processes. The 2 and 3.5% levels appear to bracket most of the coals currently in use.

Limestone Type

The limestone selected is from a quarry near the Colbert plant. Price was the main consideration; because of the very low cost at the quarry, freight is a major part of the cost delivered to the power plant.

The limestone is a high-calcium type; a typical analysis is given in Table 6. The proved reserve from which it is being quarried is large, estimated at 8 million tons of about 95% calcium plus magnesium carbonate material. Such a reserve would last some 20 to 30 years in supplying the needs of a 1300-megawatt plant such as Colbert; other undeveloped deposits in the area probably would extend the supply indefinitely.

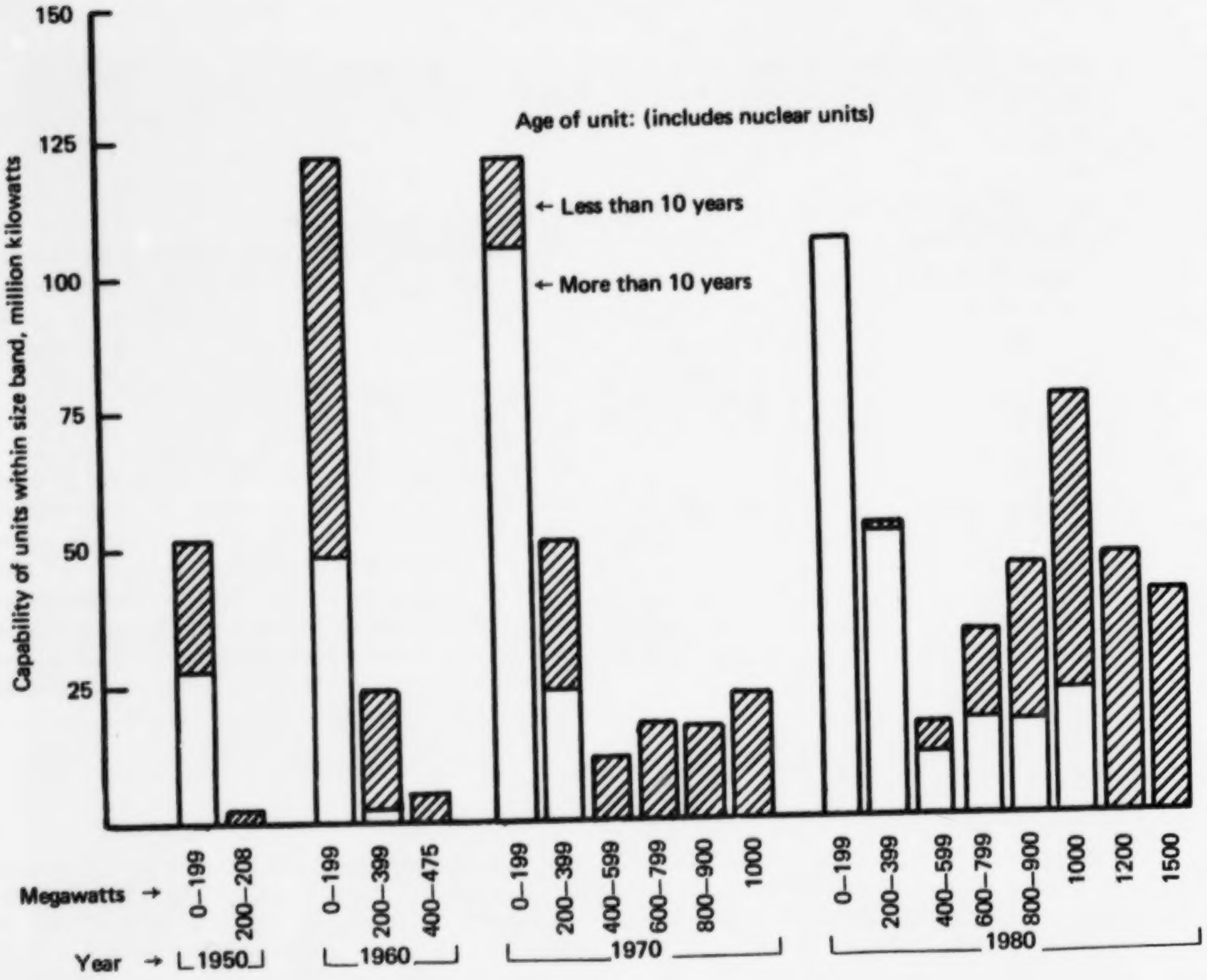


Figure 5. Distribution of thermal power plants in the United States (actual and projected)

Source: Power System Statements to FPC for 1950 and 1960

For process B, the cost of calcining the limestone before introduction into the scrubber circuit was estimated as an alternate to use of raw limestone.

Dolomite or dolomitic limestone may be a more economical choice in other parts of the country. The main difference is that pond water recirculation may be required when dolomite is used but not when high-calcium limestone is the absorbent. The cost of recycling was estimated.

Amount of Limestone Used

For the base cases, 110% of the stoichiometric requirement of limestone was assumed for process A (injection-scrubbing) and 130% for process B (scrubber addition). These levels are based on the experience of others as discussed earlier in this report. The requirement for process A is fairly well identified. The amount needed for process B is less certain and should be investigated further in experimental studies. For use of lime in process B, 100% of stoichiometric was assumed—in keeping with the British work.

Particle Size

For both processes, grinding to 70% minus 200 mesh was assumed. For injection into the boiler, this level is consistent with the experience reported and permits use of conventional coal mills. For injection into the scrubber, wet grinding was assumed as it has the advantage of eliminating the heat requirement for drying (needed with high-moisture feed to a dry-grinding operation). No other sizes were assumed as the effect of particle size on limestone utilization is not well defined. It may be that future experimental work will show that smaller particle size improves absorption. The cost of fine grinding by dry methods was discussed in the report of the dry limestone process and it is planned to evaluate wet grinding to small particle size in further studies.

Boiler Type

The boiler on which the study is based–one of the 200-megawatt units at the TVA Colbert Steam Plant (Fig. 6)–is of the pulverized-fuel type, that is, the coal is finely ground and injected through burners into the boiler. This is the most common type. The cyclone type differs in that the coal is not finely ground and is burned in cyclones (furnaces that discharge heated gas into the boiler proper through ports) set into the side of the boiler. Most of the ash in the cyclone type leaves the boiler as bottom slag, whereas most of it leaves as fly ash in the pulverized-fuel type. Thus there would be a lower total load of solids in the gas from the cyclone boiler and a much higher ratio of limestone to fly ash, which would probably introduce differences in slag formation on boiler surfaces and in removal of solids from the gas leaving the system. There is not enough information to evaluate these differences on the basis of cost.

Table 5. Sulfur Content of Coal from Various Sources[a]

Location of mine	Pyritic	Organic	Sulfate	Total
Clearfield County, Pa.	2.82	0.74	0	3.56
Franklyn County, Ill.	1.50	1.02	0	2.52
Union County, Ky.	1.05	2.23	0	3.28
Webster County, Ky.	0.70	0.78	0	1.48
Boone County, W. Va.	1.47	1.01	0	2.48
Walker County, Ala.	0.81	0.81	0	1.62
Jefferson County, Ala.	0.97	0.72	0.03	1.72
Clay County, Ind.	2.13	1.79	0	3.92
Lee County, N. C.	1.52	0.80	0	2.32
Butler County, Pa.	0.47	0.62	0.07	1.16
Allegheny County, Pa.	0.79	0.66	0.23	1.68
McDowell County, W. Va.	0.08	0.46	0.01	0.55
Letcher County, Ky.	0.13	0.51	0.04	0.68
Morgan County, Tenn.	1.75	1.78	0.71	4.24
Cherokee County, Kans.	1.99	0.71	0.32	3.02

[a] Field, J. H., Brunn, L. W., Haynes, W. P., and Benson, H. E. Bureau of Mines Report of Investigations 5469, p. 3, U. S. Department of the Interior (1959).

Table 6. Typical Analysis and Physical Properties of Colbert County Limestone

Chemical analysis, % by wt		Physical properties	
$CaCO_3$	91.8	Color	Gray
$MgCO_3$	3.1	Bulk density, lb/ft^3	150
SiO_2	3.5	Hardness (Hardgrove Index)	50
Fe_2O_3	0.3	Specific heat, Btu/lb at 100° F.	0.205
Al_2O_3	0.4		

For the boiler selected, the height is 120 feet and the cross section at the burner level 24 by 58 feet. Gas flow rate in the main portion of the boiler is about 60 feet per second; total volume of gas is about 450,000 standard cubic feet per minute. The boiler is of the balanced-draft type (fitted with both forced-draft and induced-draft fans).

Point of Limestone Addition

Injection of the limestone about 8 feet above the burners through six nozzles set into the boiler wall at one elevation and spaced across the face of the boiler (same as for the dry process) was assumed for process A. Injection with the coal might be suitable and would be simpler; it was not chosen because of possible dead-burning and consequent adverse effect on lime dissolution rate in the scrubber. A single injection point would also have simplified the installation, but it was assumed that sorption in the boiler should be optimized as much as possible without unduly increasing investment.

For process B, the lime or limestone would be added directly in the scrubber feed tank after drawing off a sidestream for control of solids concentration in the scrubber system.

Dust Collection

For an existing plant, studies indicated that the overall operating cost would be increased by continuing to operate the electrostatic precipitator. In the absence of any conclusive evidence of process advantage, it was assumed that the precipitators would be taken out of service but retained for standby service through use of bypass duct work. Normally, all dust collection would be carried out in the scrubber. If operation during extended periods without the scrubber were necessary, the precipitator could be put into operation. For new plants, elimination of the precipitator was assumed because of the large reduction in investment.

As discussed earlier, continued operation of an existing precipitator may be desirable in process A as a means of diverting lime to a delay tank in the scrubber and thereby reducing scaling. Costs were determined for this type of operation. No additional precipitator capacity was provided to compensate for any loss of efficiency as a result of reduced sulfur trioxide content of the flue gas.

Solids Disposal

It was assumed that waste solids would be conveyed to a disposal pond by sluicing rather than by solids-handling

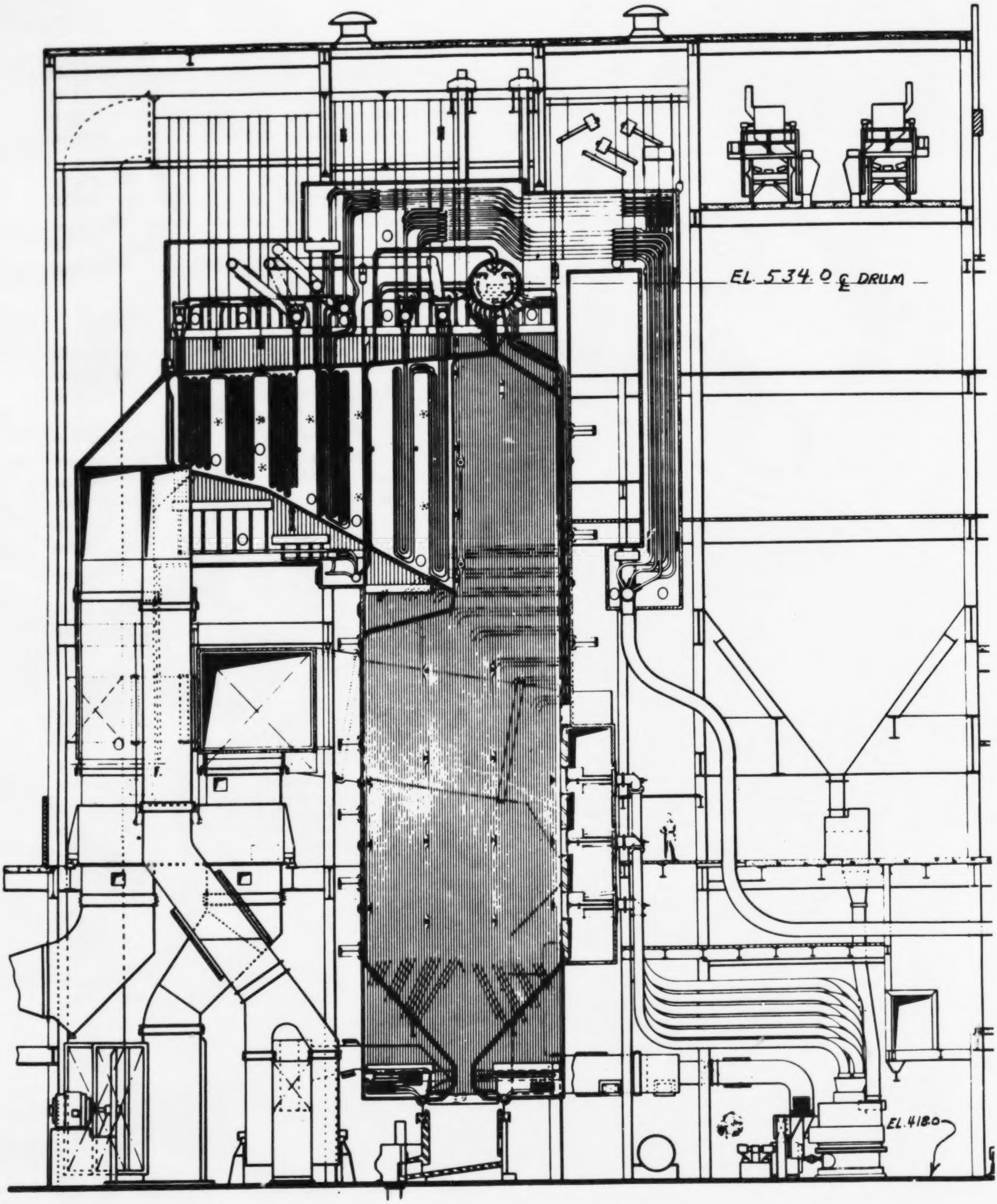

EL. 534.0 ℄ DRUM
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Figure 6. 200-megawatt unit at TVA Colbert steam plant equipment because of the known high cost of the latter. Alternates evaluated were (1) pumping the total purge stream (solids content of about 10%) to the pond and (2) concentrating the solids to a 40% slurry in a thickener before pumping. The liquid phase would be discharged to streams, the current practice in power plants for ash sluice water. Since in some situations, particularly when dolomite or dolomitic limestone is used, recycling of sluice water may be required, the cost of recycling was estimated also.

Procurement of additional space for solids storage during 10 years of operation was assumed.

Stack Gas Reheating

Reheating of the cooled stack gas to 250° F. was assumed. This level is the maximum practical for use of the waste heat in the hot gas entering the scrubber, and would approach the stack temperature from existing units equipped with low-level economizers. Costs for lower levels of reheat were also estimated.

The method of reheating was selected on the basis of cost estimates on five different methods. The most economical was use of heat exchangers to extract heat from the stack gas ahead of the scrubber and to transfer heat to the scrubber exhaust gas.

Stack Height

Removal of sulfur dioxide by wet scrubbing will greatly reduce the potential hazard of localized fumigation. It can be argued that removal of most of the sulfur oxide elminates the need for a high stack, particularly in view of the current trend to very expensive 800- to 1200-foot stacks for new large power plants. However, discussions with air quality specialists indicate that there will be considerable opposition to any major stack height reduction merely because one of the pollutants has been removed to a major extent. Other objectionable components, mainly nitrogen oxides and residual dust, are removed only to a limited degree. Moreover, during upset operating conditions or scrubber unit breakdown it would be desirable to have the high stack available for good dispersion. Therefore, no credit was taken for reduction in stack height in the basic cost estimate. As a matter of interest, stack costs are presented so that the effect on overall cost of reducing stack height can be visualized.

Equipment Selection and Description

Equipment required for the limestone–wet scrubbing process can be divided into three major categories:

1. Limestone receiving, storage, and handling.
2. Gas scrubbing and reheat.
3. Solids disposal.

The major alternatives and type of equipment selected to carry out category 1 were discussed in detail in the report of the dry process. Equipment in this category for the present conceptual design is described later in this section. There are several alternative choices in selection of equipment in categories 2 and 3.

Major Alternatives

Scrubber Type: A study of scrubber types was made, on the basis of information from plant visits, scrubber consultants, scrubber vendors, and literature sources.

The scrubber type used in the Howden-ICI pilot plant program and later at the Fulham plant was a countercurrent, grid-packed tower designed for low gas velocity (5 ft./sec.) and high-volume liquid recirculation. The pressure drop in the scrubber was only about 1 inch of water. High efficiency was obtained; when the system was operated at a pH of about 6.3, about 98% of the sulfur dioxide and dust in the inlet gas were removed.

Pilot plant studies at Wisconsin Electric were carried out in a 14-inch diameter, Turbulent Contact Absorber (TCA; moving bed of hollow plastic sphere) designed and manufactured by Universal Oil Products Company, Air Correction Division (Fig. 7). The majority of the sulfur dioxide absorption tests were made with two stages of scrubbing. The effect of liquid and gas rates on absorption efficiency with pulverized limestone (90% -200 mesh) used as the absorbent is shown in Figure 8.

Data on particulate collection efficiency in the Wisconsin Electric studies are given in Figure 9. With two-stage scrubbing and a high ratio of liquid to gas, more than 99% of the dust was removed; inlet dust loadings were in the range of 1.5 to 2.5 grains per cubic foot. Pressure drop with two stages of scrubbing was about 6 inches of water.

The scrubber design used in the Combustion Engineering – Detroit Edison work was also a mobile-bed type (Hydro-Filter, manufactured by National Dust Collector Corporation). Instead of hollow spheres, glass marbles were used in the bed. A single-stage bed gave 6-inch pressure drop and removal of about 98% of the sulfur dioxide and 99.5% of the dust.

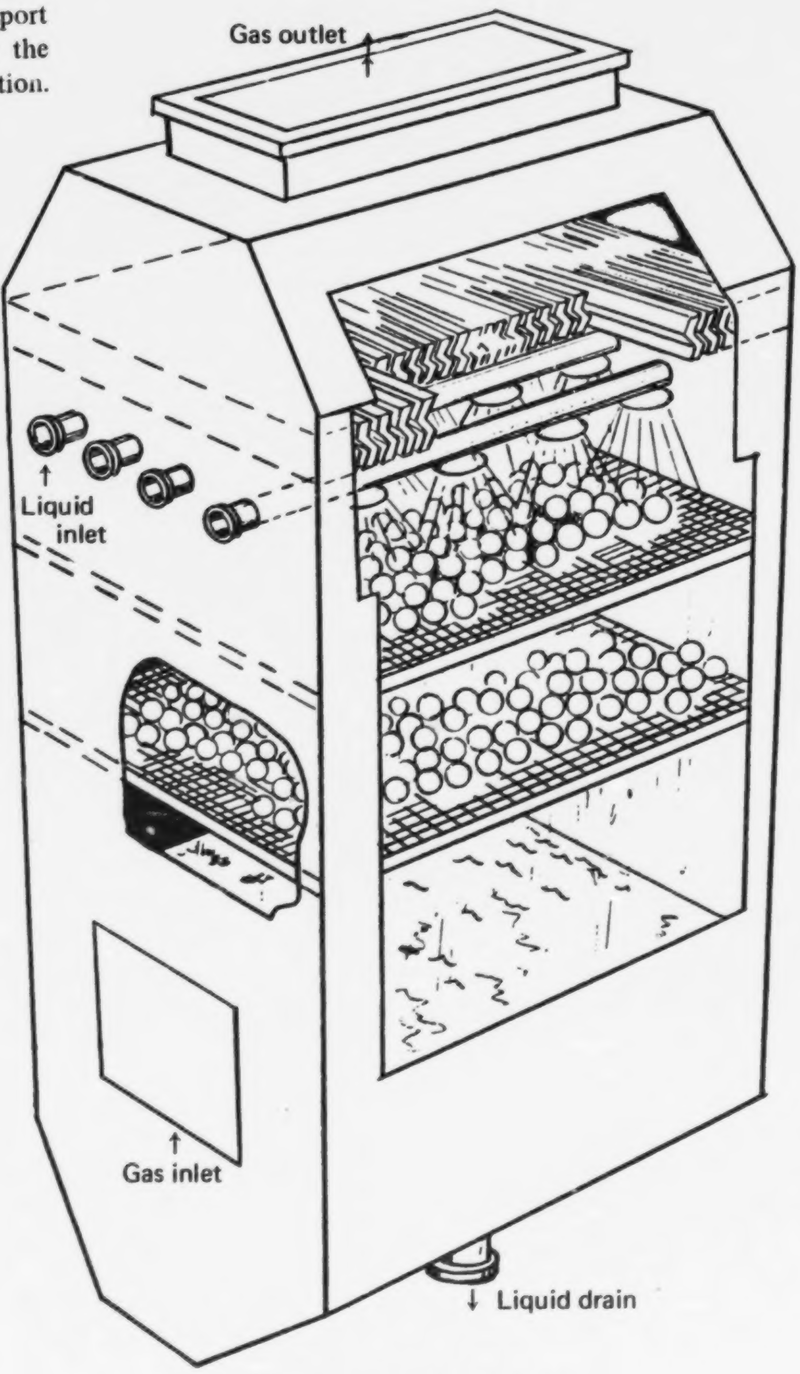
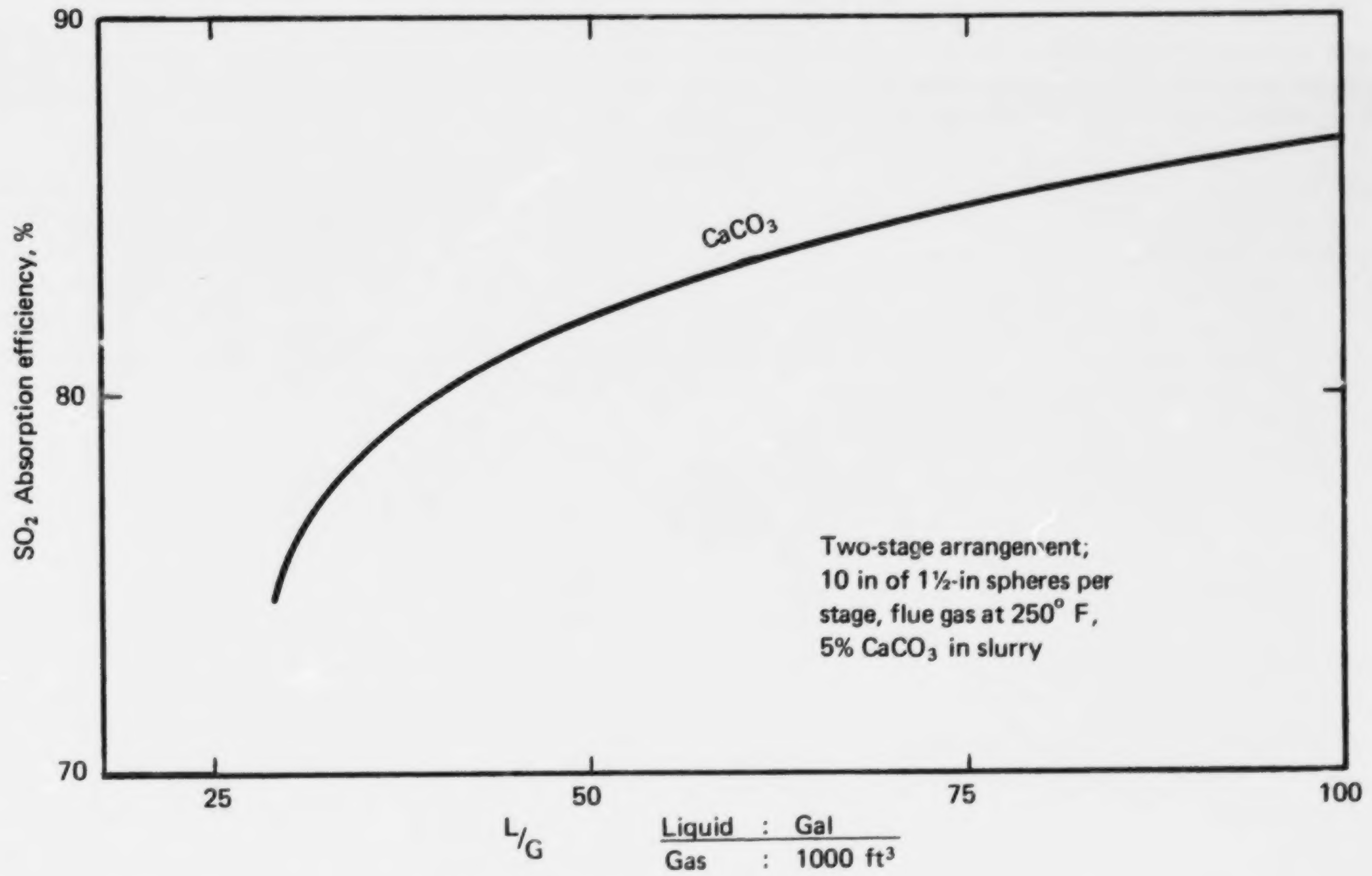
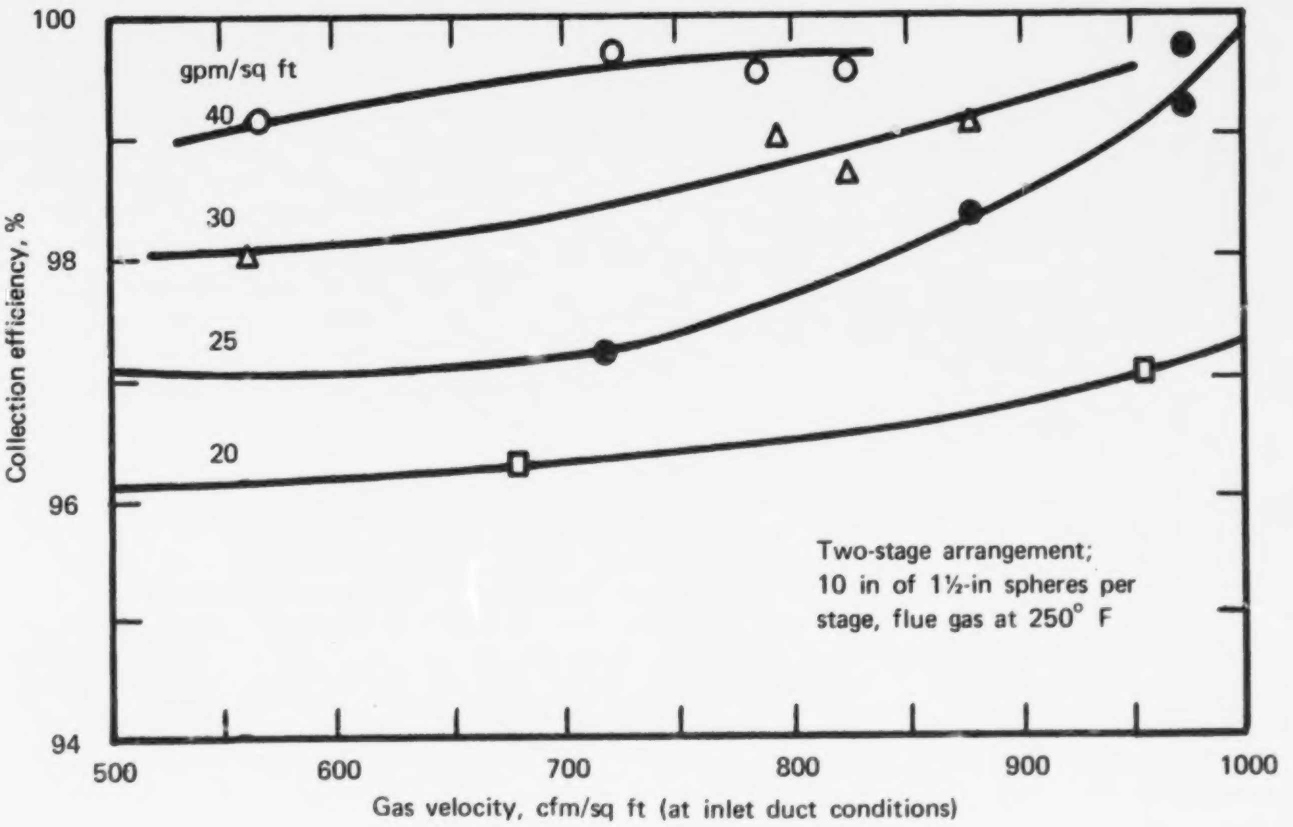


Figure 7. Typical turbulent contact absorber scrubber



Two-stage arrangement;
10 in of 1½-in spheres per
stage, flue gas at 250° F,
5% $CaCO_3$ in slurry

Figure 8. Sulfur oxide absorption efficiency in turbulent contact absorber (7)



Two-stage arrangement;
10 in of 1½-in spheres per
stage, flue gas at 250° F

Figure 9. Dust collection efficiency in turbulent contact absorber (7)

Thus only two types of scrubbers have been tested and both gave good dust and sulfur oxide removal efficiencies. The wooden grids used in the Howden-ICI work are seldom used in scrubbers today. The mobile-bed type is a fairly recent development, particularly applicable to scrubbing with a slurry because of its self-cleaning action.

Other types of scrubbers also should be applicable. In evaluating them, the following characteristics may be set down as desirable:

1. Low pressure drop (6 in. of water or less)
2. High efficiency for removal of sulfur dioxide (85-95%) and particulates (minimum 99.0%).
3. Turbulent mixing of the liquid phase and suspended solids to promote dissolution of calcium oxide or calcium carbonate.
4. Resistance to plugging by scaling or by settled solids from the circulating slurry.
5. Adequate turndown ratio (ratio of design volume rate to minimum volume rate at which scrubber will operate satisfactorily) to allow operation during reduced boiler load.

The following types of scrubbers have been considered in this study:

1. Packed[8]
    a. Countercurrent flow
    b. Cocurrent flow
    c. Crossflow
2. Sieve tray
    a. Countercurrent without downcomers
    b. Crossflow impingement plate with downcomers
3. Venturi
4. Spray
    a. Cyclonic
    b. Spray tower
5. Mobile-bed
6. Orifice

The factors that exert the most influence on scrubber design are inlet dust loading, gas volume, pressure drop, and volumetric turndown.

With limestone injection in the boiler and no dust collection ahead of the scrubber, the dust loading will be about 7 to 9 grains per cubic foot. If a dust collector is used upstream of the scrubber the inlet dust loading is reduced to about 1 grain per cubic foot or less.

In most cases, scrubber dust-removal efficiency is stated on a weight basis but the percentage of particles removed is likely to be lower. Most scrubbers, regardless of type, will effectively remove particles 10 microns in diameter or larger but removal of smaller particles is necessary for good plume appearance since about 30% of a typical fly ash is less than 10 microns in size. A 100-gram sample of this ash would have approximately the following particle distribution:

Particle size microns[a]	Weight in fly ash, g	Number of particles
0.5	0.13	8.1×10^{11}
1	0.37	2.9×10^{11}
2	1.5	1.5×10^{11}
3	2.5	7.4×10^{10}
4	3.5	4.3×10^{10}
5	1.5	9.5×10^{9}
6	3.5	12.9×10^{9}
7	3.5	8.1×10^{9}
8	4.5	7.0×10^{9}
9	4.0	4.4×10^{9}
10	5.0	4.0×10^{9}
20	20.0	2.0×10^{9}
30	12.0	3.5×10^{8}
40	8.0	1.0×10^{8}
50	8.0	5.0×10^{7}
70	9.0	1.7×10^{7}
100	3.0	2.4×10^{6}
+100[b]	10.0	2.9×10^{6}

[a] Assumes particles smaller than 0.5 micron are 0.5 micron in diameter; particles smaller than 1 micron and larger than 0.5 micron are 1 micron in diameter, etc., to 100 microns.

[b] Assumes all +100-micron particles to be 150 microns in size.

The main question in regard to this is whether a scrubber that removes the required 99% by weight will give adequate performance on the basis of number of particles removed. The difficulty is that no data of this type are available for wet scrubbing of the fly ash-lime-calcium sulfate mixture involved. Information from scrubber manufacturers indicates that various types will remove over 99% of the dust (by weight) at about 6 inches pressure drop, and from 70 to 90% of the particles (based on the distribution given above). Operation at higher pressure drop would give better removal of small particles (Fig. 10) but the cost would increase quite rapidly at high efficiencies.

High weight removal efficiency was demonstrated in the Combustion Engineering tests with a mobile-bed scrubber, in which 99.5% removal was accomplished at 6 inches pressure drop. No data are available on removal of the finer particles. However, removal higher than 99% would ordinarily be regarded as quite good without reference to number of particles. Therefore, it has been assumed that the more efficient scrubbers are acceptable at 6-inch pressure drop.

The volume of gas to be scrubbed is quite large; stack gas rates for 200-, 500-, and 1000-megawatt units are 545,000, 1,362,000, and 2,725,000 actual cubic feet per minute,

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[8] Packed with modern type of packing. The wood slat packing used in the Howden-ICI work was not considered because the surface area is low and therefore very large scrubbers would be required.

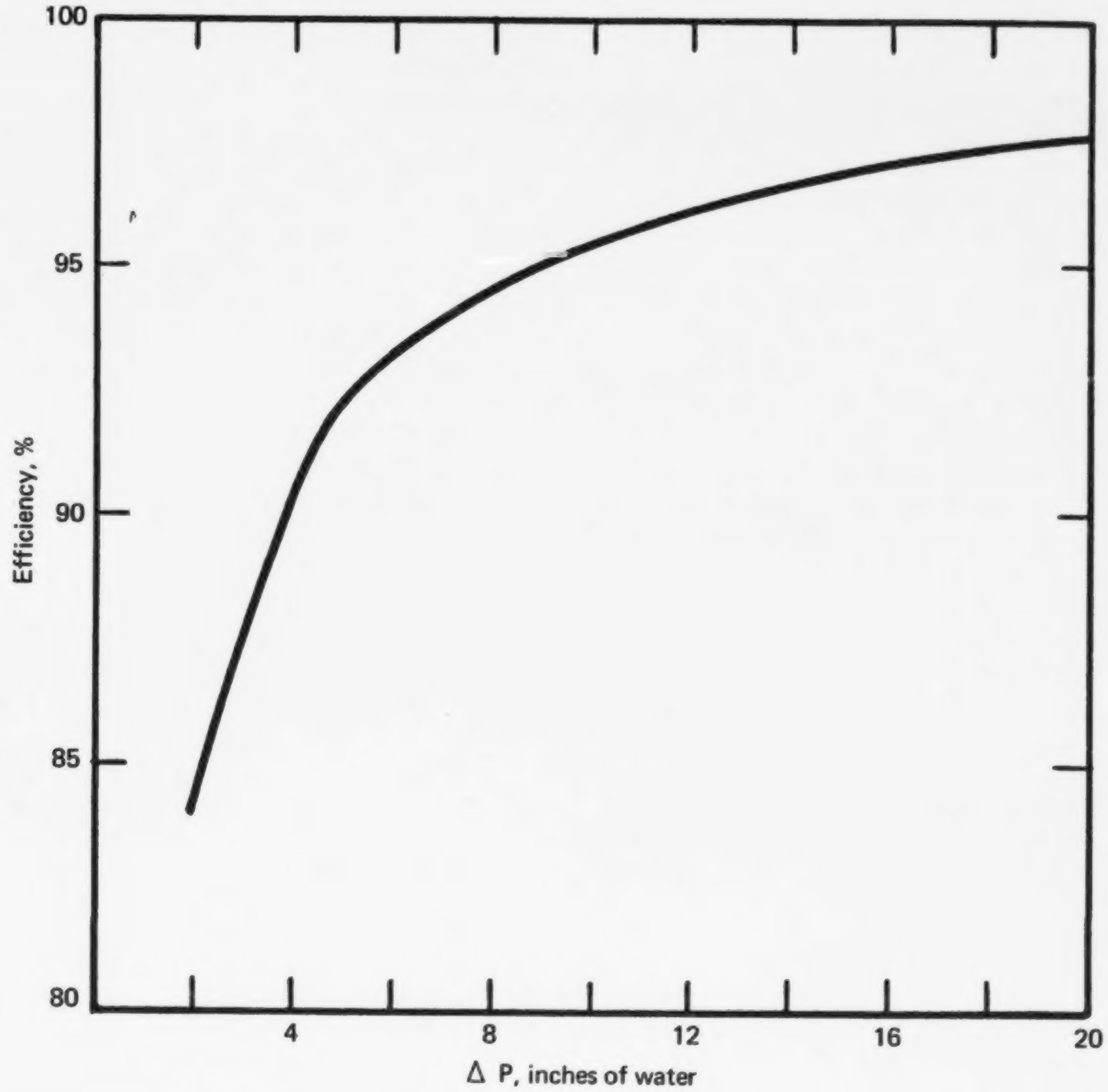

Figure 10. Collection efficiency for 1-micron particles (impingement-type scrubber)

respectively, at 130°F., the estimated scrubber exhaust temperature. To simplify gas distribution, the number of scrubbers per boiler should be as small as possible; it is assumed that not more than four would be used. Only a few manufacturers are presently capable of supplying scrubbers as large as this would require. (There is a need for pushing scrubber technology to larger sizes.) The turbulent bed scrubbers are the only type offered at this time in sizes larger than 500,000 cubic feet per minute. One manufacturer of the crossflow impingement type can build units to handle flows up to 500,000 cubic feet per minute.

Table 7 shows scrubber combinations for the three power plant sizes. The turndown ratio for the scrubbers is based on operation of the power plant at reduced load, as low as 30% of design; a typical 200-megawatt unit has six

Table 7. Comparison of Scrubber Combinations

Plant size mw	Assumed number of scrubbers	Design capacity/ scrubber ACFM	Gas flow @ max. turndown ACFM[a]	Volumetric turndown ratio required for ea scrubber in operation	Approx. number of scrubber manuf.[b]
200	1	545,000	164,000	3.33	2
200	2	272,500	164,000	1.67[c]	4
200	4	136,250	164,000	1.67[c]	6
500	2	681,250	410,000	1.67	2
500	4	340,625	410,000	1.67[c]	3
1,000	4	681,250	820,000	1.67[c]	2

[a] Assuming scrubber system would be operated at a minimum capacity of 30% of design.

[b] Presently known manufacturers for scrubbers of the sizes required.

[c] Turndown ratio required for each scrubber when using only one-half of the scrubbers at minimum system flow.

coal pulverizers and can be operated with as few as two of them, giving a load of about 60 megawatts.

A comparison of different scrubber types is given in Table 8, showing the indicated effectiveness of each for operation with or without dust collection ahead of the scrubber and with limestone either injected into the boiler or added directly into the scrubber circuit. For operation with low dust loading, packed scrubbers (both crossflow and cocurrent), cyclonic spray, and spray towers would be suitable. For use without dust removal prior to scrubbing, the mobile-bed type provides the best combination of ability to handle high dust loading, small particle removal with low pressure drop, and turbulent action which promotes dissolution of the absorbent and also prevents plugging. Either a two-stage TCA scrubber or a one-stage Hydro-Filter scrubber operating at about 6 inches of water pressure drop should remove 99% of the dust and 95% of the sulfur dioxide when limestone is injected into the boiler. Addition of limestone in the scrubber is assumed to reduce the sulfur dioxide removal efficiency to 85%; operation with a pressure drop of only 3 inches of water would further reduce the removal to 70% and dust-collection efficiency to 95%.

The mobile-bed type was selected for the economic studies. Use of two two-stage TCA units designed for 6 inches of water pressure drop was assumed for both processes A and B. Hydro-Filter scrubbers should also be acceptable; however, in the Hydro-Filter the scrubbing liquor enters at the bottom through upward sprays and is carried up through the marble bed by the momentum of the gas after which it flows back through downcomers through the bed and to the scrubber outlet. In contrast, the liquor enters at the top in the TCA scrubber and operation is fully countercurrent. This should give lower retention time and therefore possibly less scaling in the scrubber.

Another variation that might be considered is use of a two-stage scrubbing system as a means of avoiding introduction of the lime directly into the scrubber (in process A) and thereby reducing scaling. The first stage would be designed primarily to remove the solids, which would then be introduced into the recirculating liquor going to the second stage. For example, a venturi scrubber

Table 8. Scrubber Comparison

Scrubber type	Usual application	Approx. max. *inlet* gas dust loading grains/ft³	Approx. dust-removal efficiencies[a] 1-micron particles	2-micron particles	5-micron particles	Expected pressure drop for 95% SO_2 removal, in. water	Dissolution of reactants	Resistance to solids plugging[b]	Approx. turndown ratio[c]
Countercurrent packed[d]	GA	1-2	0	0	95	8	G	NR	1.2
Cocurrent packed[d]	GA	2	0	0	95	6	G	UC	5.0
Crossflow packed[d]	GA	3-4	0	0	95	4	G	UC	5.0
Countercurrent sieve	GA & PR	10	85	92	95	6	G	UC	1.2
Crossflow sieve	GA & PR	100	93	96	98-99	6	G	NR	1.2
SF venturi	GA & PR[a]	+100	97	99	99+	12	NR	E	1.3
Splash plate venturi	GA & PR[a]	+100	97	99	99+	12	UC	E	2.0
Cyclonic spray	GA & PR[a]	15-20	30	70	94	4-6	F	E[e]	2.0
Spray tower	GA & PR[f]	15	20	60	90	3-5	F	E[e]	4.0
Mobile bed, Hydro-filter	GA & PR	25	92	98	99+	6½	E	E[e]	1.5
Mobile bed, TCA	GA & PR	25	92	98	99+	6-7	E	E[e]	1.5
Orifice	PR	25	35	75	93	6-8	NR	E	1.2

[a] When operating at pressure drop required for 95% sulfur dioxide removal.
[b] For a system in which calcium compounds are precipitating.
[c] Ratio of design volume rate to minimum allowable volume rate.
[d] These scrubbers will not remove particles smaller than 3 microns unless nucleation is effected by condensing water vapor.
[e] Provided spray nozzles do not plug.
[f] For highly soluble gases only.

Key: E–excellent; G–good; F–fair; NR–not recommended; GA–gas absorption; PR–particulate removal; Y–yes; N–no; UC–uncertain.

might be used for the first stage; it would have good dust-collection efficiency but would not be expected to dissolve the lime very well, which would minimize calcium salt precipitation in the unit. The collected ash-lime mixture would then be introduced into the slurry circulating through a second scrubber designed to dissolve the absorbent and with adequate retention time to absorb the sulfur oxides. This unit could be of the spray tower of mobile-bed type.

No data are available for evaluating this arrangement.

Mist Eliminators: The removal of liquid entrained in the scrubber exit gas serves three purposes:

1. Reduces the load on the stack gas reheater.
2. Decreases the deposition of liquid in the fan and in ducts located downstream from the scrubber.
3. Reduces amount of solids discharge by removing the dissolved solids contained in the mist which, after reheating, would be emitted to the atmosphere as dust.

The mist emitted by a wet scrubber would be comprised of liquid particles ranging from about 10 microns to 60 microns in diameter; particles smaller than 10 microns would not normally be produced unless water were condensed from the flue gas. With no mist removal from the TCA scrubber exhaust, the gas would contain about 0.012 pound of entrained liquid per cubic foot of gas. Use of a 95% efficient demister would reduce the heat required for reheat by about 3 million B.t.u. per hour or about 4% of the total required for reheat to 250°F.

The mist eliminator should have the following characteristics:

1. A removal efficiency of at least 70% for 10-micron liquid particles and an overall efficiency of about 95%; above this the high capital cost is not justified by the advantages.
2. The ability to flush (with the collected mist) any undissolved solids collected in the separator.

Several types of mist eliminators were evaluated. The TCA and Hydro-Filter scrubbers both are equipped with an impingement vane type of mist eliminator. The shape of the vanes and their arrangement cause impaction and coalescence of the mist. Other types of entrainment separators evaluated were the swirl vane, wire mesh (York), fiber bed (Brink), cyclonic, and packed bed (6 to 12 in. of Tellerettes, Pall Rings, or others).

A comparison of the mist eliminators is shown in Table 9. From this it appears that the vane types offer the lowest investment and best plugging resistance at acceptable pressure drops. The overall removal efficiency of the vane types should be about 95%. The Brink type of mist eliminator is more efficient than justified for this application. The wire mesh and packed-bed entrainment separators could be used but should be equipped with a means for flushing with fresh water to remove solids collected on the mesh or packing. The cyclonic type (swirl chamber or tangential scrubber outlet) would probably not give an overall efficiency of 95%.

An impingement vane mist eliminator was selected for use with the TCA scrubber in the conceptual design.

Equipment for Reheating Gas: Alternate methods studied for reheat of the stack gas were:

1. Install a combustion system at the base of the power plant stack for burning natural gas, oil, or coal and mix the combustion products with the scrubber exit gas.

The main advantages for this method are low investment, flexibility in degree of reheat, minimum added pressure drop, low maintenance, and good reliability. Disadvantages are fuel cost, introduction of objectionable components (SO_3, SO_2, ash) into the gas, and fuel supply problems. Natural gas is the most expensive fuel but the added cost might be justified on the basis of convenience and clean combustion. However, gas would not be available at many power plants. Oil would be less expensive but receiving, storing, and handling would be a problem. Use of coal would result in the lowest fuel cost but would add more sulfur

Table 9. Comparison of Demisters

Type	Removal efficiency for 10-micron mist particles, %	Pressure drop, in. of water[a]	Approx. cost for 200-mw installation, $	Resistance to solids plugging
Impingement vane	70	0.4	3,500	Good
Swirl vane	70	0.3	3,000	Good
Wire mesh (York)[b]	98-99	0.8	12,000	Fair
Fiber bed (Brink)	100	7.0	75,000	Poor
Packed bed[c]	90	0.1-0.2	9,000	Fair
Cyclonic	60-70	0.5-1.0	5,000	Good

[a] Required for removal efficiency shown for 10-micron particles.
[b] Six-inch-thick bed.
[c] Eight inches of packing.

dioxide and ash to the stack gas. The fly ash emission could be minimized by firing the coal on a grate stoker.

2. Bypass the scrubber with part of the gas stream and mix this gas with the scrubber exit gas.

This procedure requires minimum investment and has essentially no operating cost. As discussed earlier, however, bypass of the 310° F. gas from a point after the air heater increases ground-level concentration rather than decreasing it. Bypass of gas from upstream of the air heater would be more effective but this would reduce boiler efficiency and possibly give rise to problems in removing the dust.

3. Use heat exchangers for direct transfer of heat from the scrubber inlet gas to the exhaust gas.

With this method, heat that would be wasted is recovered. Further advantages are reduction in the amount of water required for evaporative cooling, a corresponding reduction in gas volume, and, except for maintenance, no labor requirement. Disadvantages are the large heat exchanger required (because of low transfer coefficient and temperature differential), high pressure drop, and possibility of fouling–which would lead to low efficiency and high maintenance cost. For addition of limestone in the scrubber circuit, corrosion by sulfur trioxide would be a problem with mild steel exchangers.

4. Use a cyclic-liquid heat exchange system with heat transfer from the inlet gas to treated water and from the water to the scrubber exhaust gas.

The better heat transfer coefficient would permit use of smaller exchangers than those required for gas-gas exchange and the smaller surface would reduce pressure drop and maintenance.

5. Heat with steam from the turbine cycle in a heat exchanger at the scrubber outlet.

This method would require additional coal firing in the boiler to generate the extra steam and modification of the turbine to allow higher than normal extraction rates. Extensive modification of existing units would be impractical, but in a new plant a system could be included to provide the steam required.

Use of steam for reheat would require relatively small heat exchangers installed only on the scrubber discharge, where the gas is relatively clean. Corrosion, fouling, and pressure drop would be minimized. The main disadvantage is the added fuel requirement.

6. Use a cyclic system comprised of heat exchange towers where liquid or solid particles are sprayed into the gas stream ahead of the scrubber and the sensible heat gained by the particles is transferred to the scrubber exit gas in a similar chamber.

With this system, there would be no heat exchangers to foul or corrode and the pressure drop would be low. With a liquid, partial dust removal could be effected by filtering or centrifuging the liquid from the "hot" tower. However, a low vapor pressure over the liquid would be required to prevent carryover to the scrubber and the liquid should be nonflammable or have a high kindling temperature to prevent fire hazard. Use of solid particles would require a material with good abrasion resistance to withstand the rough handling.

Insufficient information is available to prepare a cost estimate on this reheat method.

The cyclic-liquid heat exchange method of reheat was chosen for the conceptual design; a cost estimate (see economics section) showed it to have the best combination of investment and operating costs.

With process B, the gas entering the scrubber will contain sulfur trioxide whereas with limestone injection the sulfur trioxide is reacted with lime in the boiler. The portion of the heat exchanger at the scrubber inlet exposed to gas at a temperature below the dew point of sulfur trioxide (about two-thirds of the surface) was assumed to be constructed from a corrosion-resistant alloy steel. Addition of sufficient limestone in the boiler to react with the sulfur trioxide would probably prevent the corrosion problem but adding absorbent at two points would complicate the process.

Equipment Description

The plant arrangement for receiving and handling limestone is the same as presented in the report of the dry process. Limestone is received by truck shipment, dumped into an unloading hopper, and conveyed to the storage silo. For process A, the material is conveyed from storage to a surge hopper over the mill, ground, blown by the mill-sweeping air to a feed bin equipped with a bag filter vent, passed through a variable-speed screw feeder, fed into the air injection pipe by a rotary valve, and collected along with fly ash in the scrubber (Appendix D, Fig. D-1 and D-2). For process B, the limestone is conveyed from storage to a surge bin, weighed, ground in a wet ball mill, pumped as a slurry to a hold tank, and metered into the scrubber recirculation tank (Appendix D, Fig. D-3). The equipment items described below are for process A (Appendix D, Fig. D-4). The same equipment, except for slight differences in size, is used for process B except where indicated (Appendix D, Fig. D-5).

Unloading Hopper: The unloading hopper is a 20- by 20-foot concrete hopper with sides sloping to a center opening above a belt conveyor 12 feet below grade; the top of the hopper is at grade for dump truck receipts. The hopper will accommodate about 50 tons of limestone or about three truckloads.

Unloading Conveyor: The unloading conveyor transports stone from the unloading hopper to a storage silo. The conveyor is a truss-frame, troughed-belt type 160 feet in length with a 24-inch belt.

Storage Silo: Incoming limestone is stored in a steel silo 25 feet in diameter with 30-foot straight sides. The silo will accommodate about 600 tons of limestone or about a 60-hour supply for 110% stoichiometric treatment, assuming 3.5% sulfur in the coal; this reserve is sufficient for operation over a weekend without delivery. The storage silo discharges onto a transfer conveyor belt.

Transfer Belt: Limestone is conveyed from the storage silo to a surge hopper by a 230-foot belt of construction identical to the unloading conveyor.

In process B, limestone is discharged from the storage silo onto an 80-foot-long belt conveyor which empties into a 20-foot high, continuous bucket elevator with a capacity of 15 tons per hour. The elevator discharges into a surge hopper above the mill feeder; the hopper and feeder are similar to those described for process A. The feeder discharges into another elevator (same specifications) which empties into the mill.

Surge Hopper: A surge hopper is provided above the mill to protect the injection system from minor delays in operation of the transfer system. The hopper is a mild steel tank 12 feet in diameter with 15-foot straight sides and a conical bottom. The storage capacity is equivalent to 8 hours' supply.

Limestone Feeder: Rate of feeding limestone to the mill is controlled by a belt-type feeder with a capacity of 15 tons per hour.

Pulverizer: The pulverizer, which reduces the limestone from minus 1/4 inch to 70% minus 200 mesh, is a B and W type E bowl mill of the type used for grinding coal. The mill has a capacity of 15 tons per hour when grinding coal to 70% minus 200 mesh. Air for classifying and drying in the mill and transporting pulverized limestone to the feed hopper is pulled from the duct that supplies preheated air to the coal burners.

For process B, a Denver ball mill, 7-foot diameter by 10 feet long (measured inside the wear plate liners) and designed to operate in closed circuit with a 60-inch diameter Model 150 Denver spiral classifier, was selected to wet grind the required 10 tons of limestone per hour from 1/2 by 0 inch to 70% minus 200 mesh. The mill is equipped with a 300-horsepower motor. With a larger drive (350 hp.) and larger classifier (72-in. diameter), the mill could be used to produce a finer product, about 95% minus 325 mesh, with less than 10% increase in operating costs.

Limestone is fed continuously to the mill along with recycled oversize; makeup water is added in the mill. The mill discharge slurry flows to a spiral classifier where coarse particles settle and are recycled to the mill. The onsize particles do not settle quickly enough to be picked up by the spiral and are carried out of the classifier in the liquid overflow as a 10% slurry. The recycle to product ratio is about 7:1.

Ground Limestone Feed Tank: The ground limestone feed tank serves both as a receiver for the pulverizer and a supply for the injection system. The tank is a mild steel, cone-bottomed unit 8 feet in diameter and 10 feet high equipped with a bin-vent type bag filter capable of handling 300 cubic feet per minute of air. An exhaust fan is provided to compensate for the pressure drop across the filter. The tank is mounted on load cells for rate checks.

For process B, a 150,000-gallon tank is provided to store ground limestone as a 10% slurry ready for addition to the scrubber recirculation tanks. The storage tank, equipped with an agitator, will accommodate an 8-hour supply of limestone to allow for minor maintenance of the mill.

Injection System: Ground limestone is metered into the injection system through a 6-inch screw feeder equipped with a variable-speed drive; capacity of the injection system is 15 tons per hour. The feeder discharges through a 6-inch rotary valve into a mixing nozzle. This single metering system supplies absorbent for all the coal burned; instrumentation for ratio control is provided. The limestone is transported in dense phase (0.07-0.08 lb. air/lb. of solids) to the injection point through a flow splitter, a cylindrical tank with six 2-inch pipe outlets equipped with orifice plates. The flow of air and entrained solids is divided into six streams and conveyed through separate 2-inch pipes to flow nozzles that project through the boiler wall; the nozzles can be pivoted in a vertical plane to vary the angle of injection. Separate rotary compressors (maximum of 10 p.s.i.g. discharge pressure) will provide up to 400 cubic feet per minute each for conveying and injecting the limestone; the injection air will be added to the transport stream in mixing tubes just ahead of each nozzle.

Duct Modification: Existing mechanical dust collectors are removed and new ducts (8 by 13 ft.; insulated mild steel) provided to direct the gas flow from the two air heaters to the scrubbers, bypassing the precipitator. Blanks inserted at flanges on the precipitator inlet would have to be removed with rigging equipment if activation of the precipitators were needed. Discharge ducts (7 by 12 ft.) are provided to return the gas to the existing precipitator discharge duct. The general arrangement of ducts and scrubbers is shown in Appendix D (Figs. D-6 and D-7).

Scrubbers: At full load, each of the two scrubbers will handle 272,500 actual cubic feet of gas. The units are two-stage, countercurrent, Turbulent Contact Absorbers manufactured by the Air Correction Division of Universal Oil Products Company (Appendix D, Fig. D-8). The scrubber shell has a 15- by 20-foot rectangular cross section, is 40 feet high, and is constructed of mild steel lined with a 1/8-inch bitumastic coating. Internal support grids and spray nozzles (for liquid distribution) are of stainless steel construction. The bottom of the vessel serves as the receiver for the scrubber liquor, which drains by gravity flow into a recirculation tank. The mobile bed is composed of 1-1/2-inch-diameter polypropylene spheres. The scrubbers are designed for 6-inch pressure drop with an L/G volume ratio of 50.

At reduced load, one of the scrubbers will be shut down by closing louvered-type dampers installed at the scrubber inlet.

Separate recirculation tanks are provided for each scrubber. Each has a capacity of 45,000 gallons, is 22 feet in diameter by 14 feet high, and is equipped with a cone bottom for discharge of solids from the system through a variable-speed pump. Separate circulating pumps deliver 8,000 gallons per minute at 110-foot head with 300-horsepower drives; installed spares are provided. The system is designed for 10% solids in the slurry. A pH monitor is installed in the circulation tanks.

Reheat System: Mild steel, finned-tube heat exchangers are installed in the duct work on each side of the scrubber. The tubes have an outside diameter of 2 inches and 3/4-inch fins spaced five per inch. Based on a coefficient of 8 B.t.u. per hour per square foot, 145,000 square feet of surface area is required. The tubes are 20 feet long and are arranged in 8- by 16-foot tube bundles.

Circulating water (treated boiler water) is used to transport heat from one exchanger to the other. In a cold climate, antifreeze would need to be added for periods when the system was shut down, such as in peak-load operation. The design temperature profile is:

Heat removal, ° F.	
Gas	300-160
Liquid	139-275
Reheat, ° F.	
Gas	118-250
Liquid	275-139

Liquid recirculation rate is 600 gallons per minute.

For process B, approximately 100,000 square feet of the exchanger at the scrubber inlet is constructed from Corten, a steel alloy containing nickel, chromium, and manganese.

Induced-Draft Fans: The overall increase in pressure drop is 12 inches of water (6 in. in scrubber, 4 in. in reheat system, and 2 in. in duct work). Existing fans (1200-hp. motors) on both trains of the exhaust system are replaced with new fans (1500-hp. motors) designed for 26-inch static pressure. In some installations the required static pressure could probably be met by equipping existing fans with larger, higher speed motors. In others, installation of a modified impeller in an existing fan housing, in addition to a drive change, might suffice.

Waste Disposal: Rate of scrubber effluent discharge will be controlled with a variable-speed pump connected to the conical bottom of each recirculation tank. The pumps discharge into a surge tank which serves as a supply for a 1000-gallon-per-minute slurry pump. Slurry is pumped 1500 feet through a new 12-inch line to the storage pond area. Water flush connections are provided at the slurry line inlet for use if solids settling is a problem.

Storage Pond: The impounded volume for disposal of ash was increased to accommodate the additional solids load resulting from limestone scrubbing. It was assumed that acreage would be available at a price consistent with current land value. For continuous operation of the scrubber system with gas from combustion of coal containing 3.5% sulfur, about 30 additional acre-feet will be required annually. Assuming impoundment to a height of 10 feet, which is current practice for fly ash at Colbert, 3 acres more per year are needed. Cost of pond development for 10 years of operation is included in the investment estimate.

Economic Evaluation

The major cost factors in the limestone–wet scrubbing process are examined in this section. Included are evaluation of process alternatives, raw material purchase and shipping costs, and overall process investment requirements and operating costs. Representative curves are included where applicable and detailed cost estimates are given in Appendix C.

Basis of Evaluation

The assumptions and design criteria discussed previously form the basis for evaluating the two processes under consideration–limestone injection followed by wet scrubbing (process A) and lime or limestone addition directly into the scrubber circuit (process B). To get a uniform basis for comparing these, estimates were first made to determine the most economical of the several alternatives in regard to number of scrubber stages, type of solids disposal, limestone versus lime for process B, precipitator-scrubber operation, and stack gas reheat. Also examined were the cost of system pressure drop and stack gas reheat temperature so that their effect could be identified. Finally, after making the basic comparison, the effects on overall process economics were determined for such variables as unit size, sulfur content of coal, onstream time, and raw material price.

The operating costs reported include appropriate capital charges and overheads and the investments include engineering costs, contractor fees, and contingency charges. Since operating cost is handled in this way, differences in investment between process alternatives are given proper weight.

Fixed Capital Charges: Suggested guidelines for applying fixed charges to cost evaluations of private and public power production and distribution systems are outlined by the Federal Power Commission (FPC) (20). A further treatment of the economics of the regulated power industry is given by Phillips (21).

In the FPC report, the components of fixed charges are given as cost of capital, depreciation, interim replacements of equipment having short life, insurance, and taxes. These are applied as a percentage of power plant investment. The term "cost of capital" is defined as the annual percent of investment which a company should receive to maintain its credit, pay a return to the owners, and ensure attraction of money for future needs. As used in the FPC formula for fixed charges, cost of capital includes the interest on bond debt and a return on equity investment.

The fixed charges applied throughout this evaluation were adapted from the guidelines suggested by FPC; the investment percentages are shown below:

	Annual % of fixed investment	
	Existing unit	New unit
Depreciation (average of 20-yr life assumed for existing units; 35-yr life for new power unit), straight line method	5.00	2.85
Interim replacements (equipment having between 20- and 35-yr life)	–	0.65
Insurance	0.25	0.25
Cost of capital (capital structure assumed to be 50% debt and 50% equity)		
Bonds at 6% of average undepreciated investment	1.50	1.50
Equity at 11% of average undepreciated investment	2.75	2.75
Taxes		
Federal	2.80	2.80
State	2.20	2.20
Total fixed charges	14.50	13.00

Credits for Reduction in Investment and Operating Cost in Parts of the Power Plant Other Than the Scrubber: In most existing fossil fuel-fired power plants the particulate matter in the stack gas is removed by mechanical dust collectors and/or electrostatic precipitators and the gaseous contaminants are dispersed into the atmosphere by high stacks. In this evaluation it is assumed that existing power units are equipped with electrostatic precipitators (about 90% efficiency) and 300-foot stacks. If a new power unit were to be built, it would in all probability have a more efficient electrostatic precipitator (probably at least 99%) and a higher stack.

When a limestone-wet scrubbing process is added to a power unit, other types of dust collectors are not required. In the case of existing units equipped with an electrostatic precipitator, an operating cost credit could be claimed if the precipitator were not operated. For new power units, both an operating credit and an investment credit would be valid if a precipitator were not installed. Such credits are taken in this evaluation where appropriate.

An investment credit may or may not be justified for reduction in stack height. The cost per foot of such stacks increases rapidly with height (Fig. 11), and a limited reduction in height would give a disproportionately large reduction in investment. As noted earlier, it has been assumed that no credit for lower stack investment would normally be justified. In cases where justification exists, the appropriate investment to be credited can be obtained from Figure 11. Although this curve does not show directly the effect of unit size, size is included, in effect, because the curve is based on actual costs of a variety of units in which stack height increased as the unit size increased.

Credit has also been taken for reduced corrosion of boiler heat-exchanger surfaces when process A is used. The corrosion reduction credit is applied to maintenance cost only since it is assumed that the amortization life of the power plant will remain unchanged.

Process Alternatives

To provide a basis for overall evaluation, the following process alternatives were evaluated:

1. *Number of Scrubber Stages:* Both one- and two-stage scrubber units, with each stage designed for a pressure drop of approximately 3 inches (water), have been considered. Because of the higher pressure drop, the two-stage scrubber will remove more of the dust and sulfur dioxide from the stack gas; however, the investment and operating cost will be greater than for a single-stage unit.

The two-stage unit was chosen because of its higher gas-scrubbing efficiency. It should be recognized, however, that applications will probably exist for single-stage scrubbing, e.g., power plants in which more than 3 inches of added pressure drop would place an intolerable operating or economic burden, or situations where the higher degree of sulfur dioxide and dust removal is not required. An economic analysis of the investment and operating costs for both one- and two-stage units in a 200-megawatt existing power facility is given in Table 10 along with expected dust and sulfur dioxide removal efficiencies. In addition, the cost of system pressure drop over the range of 14 to 32 inches water is shown graphically in Figure 12.

2. *Solids Disposal:* Current methods of solids disposal at coal-burning power plants range from sluicing to nearby ash ponds to dry transport, sometimes for long distances, by rail, truck, or barge. The operating costs

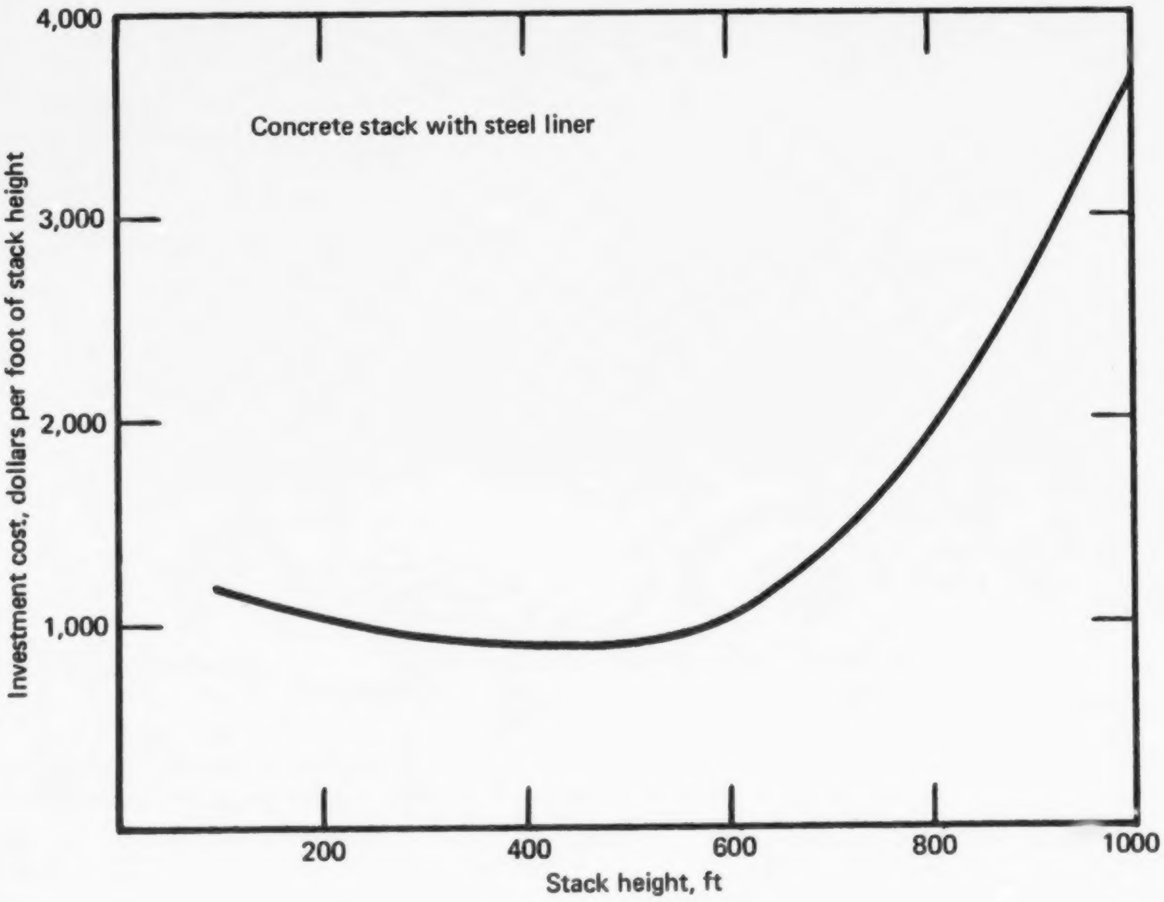

Figure 11. Cost of power plant stacks

associated with these methods range from $0.50 to $4.00 per ton of solids or higher. Since the limestone–wet scrubbing process more than doubles the quantity of solids to be disposed of (from 52,000 to about 157,000 tons/yr. from a 200-mw. unit burning pulverized coal containing 12% ash and 3.5% S and with 110% stoichiometric $CaCO_3$ injection), its use would be most applicable to those plants that have low-cost methods of solids disposal. Generally, these would be plants having on-site disposal ponds and available additional land area. A plant of this type was assumed for the evaluation.

Table 10. Investment and Operating Cost for One- and Two-Stage TCA Scrubber Systems in Existing 200-mw Unit

Process	Scrubber stages	Dust removal efficiency %	SO_2 removal efficiency %	Oper. cost of scrubber system/yr[a] $	Investment for scrubber system[b] $
A	One	98.0	85	232,000	807,000
A	Two	99.5	95	248,800	846,000
B	One	98.0	70	234,300	813,000
B	Two	99.5	85	250,100	853,000

[a] Operating cost includes power, labor, maintenance, overheads, and 14.5% fixed charges on the scrubber, hold tank, circulating pumps, and additional exhaust fan and motor capacity required because of higher pressure drop in the system.

[b] Investment includes installed cost of scrubber, hold tank, circulating pumps, and additional exhaust fan and motor capacity required.

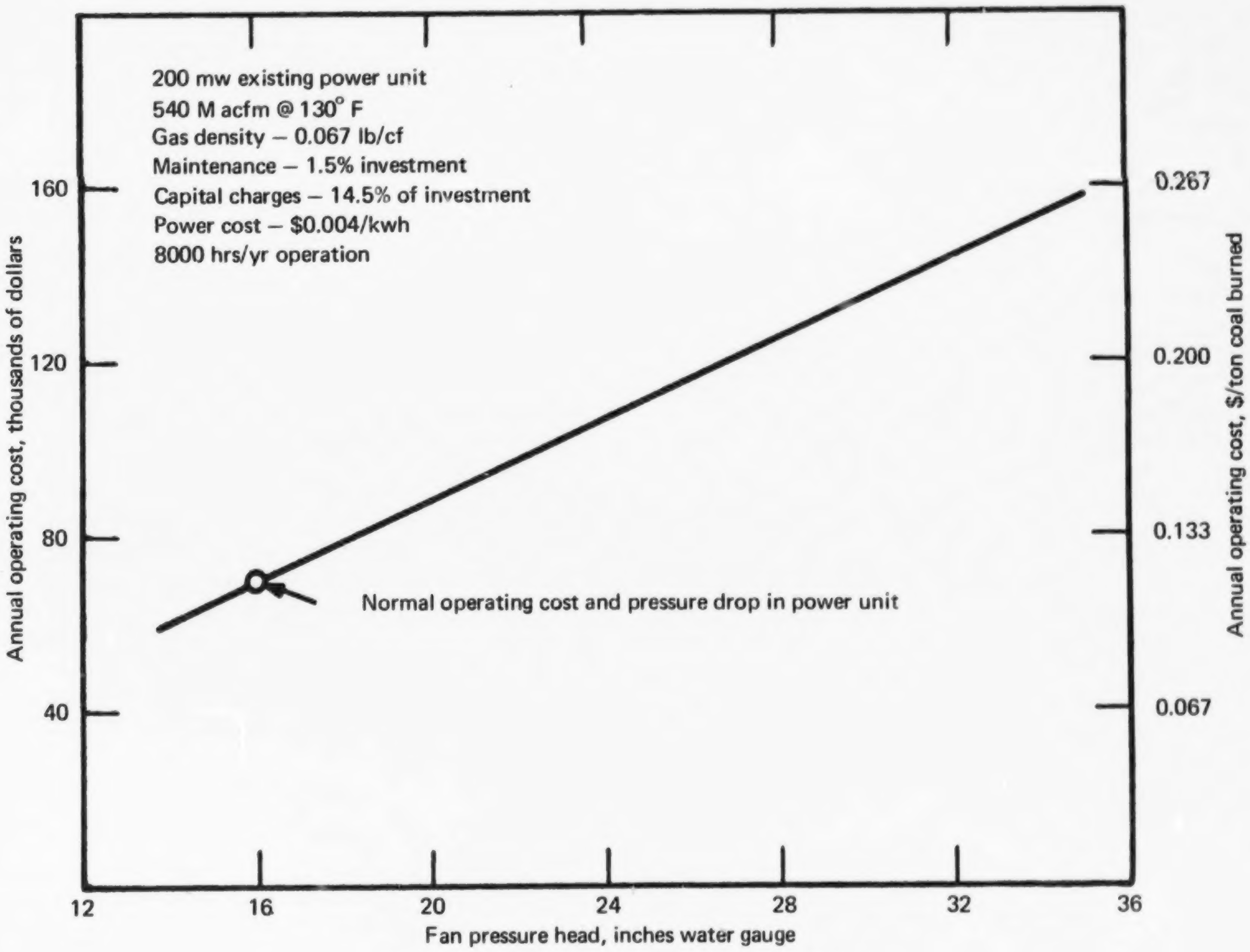

Figure 12. Effect of pressure drop on operating cost of stack gas exhaust fan

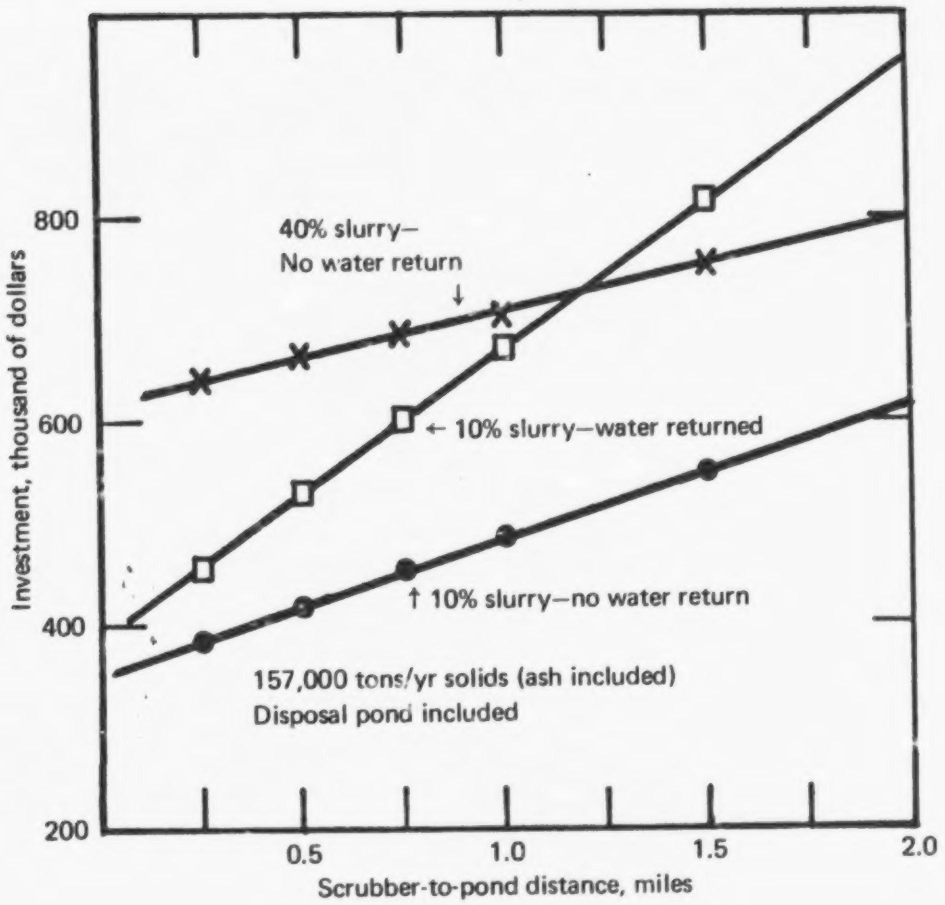

Figure 13. Investment for various methods of solids disposal

Three methods of on-site solids disposal were compared for various scrubber-to-pond distances. These were (1) direct pumping from scrubber to pond of dilute waste slurry (10% solids) with no return of water, (2) direct pumping to pond of 10% slurry with the pond serving as a settling basin before return of the water for reuse, and (3) thickening to a 40% slurry and pumping to the waste pond with no return of water. The costs shown are based on the amount of solids that would be produced by a 200-megawatt unit using process A with coal containing 3.5% sulfur and 12% ash. A value of $0.10 per 1000 gallons is placed on the water.

The results are shown in Figures 13 and 14. Method 1, direct pumping of 10% slurry with no return of water, requires the least investment, has the lowest operating cost for distances over 0.6 mile, and, incidentally, is the method used for handling fly ash in most coal-burning power plants. It was selected as the method to be used in the further estimates.

3. *Use of Calcined Limestone in Process B:* If calcined limestone were used in place of raw limestone in process B. more efficient scrubbing would be realized with a resultant reduction in material requirements and solids-handling and disposal cost. However, based on the cost of calcining determined in the dry process study, the higher price of calcined limestone more than offsets any savings due to higher scrubbing efficiency (Table 11).

4. *Precipitator-Scrubber Operation:* Addition of a wet-scrubbing process to a power plant already equipped with an electrostatic precipitator requires a choice between operating and bypassing the precipitator. Since process operating and investment costs are affected, an appraisal of the alternatives was made with the results shown in Table 12.

The results indicate that operating the precipitator would increase costs for process A and would offer little advantage in process B. Therefore further estimates for existing power units are based on inactivating the precipitators.

5. *Stack Gas Reheat:* Five different methods (see earlier description) of reheating the scrubbed and cooled stack gas were evaluated: (1) direct heating with natural gas, (2) indirect heat exchange between the stack gas and a heat-transfer liquid, (3) indirect heat exchange

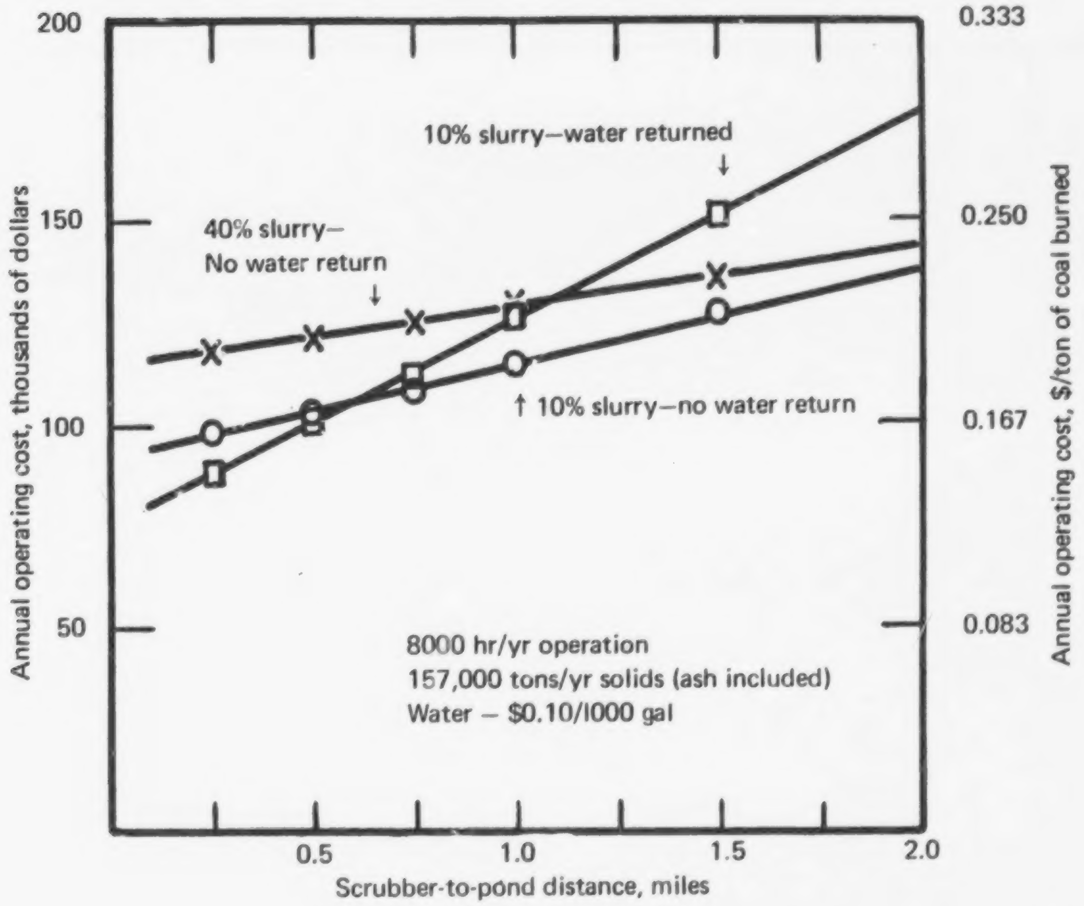

Figure 14. Operating cost for various methods of solids disposal

Table 11. Effect of Absorbent Type on Operating Cost for Process B[a]

	Raw limestone, 130% stoichiometric		Calcined limestone,[b] 100% stoichiometric	
	200-mw	1,000-mw	200-mw	1,000-mw
Limestone, tons/yr	96,000	480,000	75,000	375,000
Annual operating cost, $				
Raw material	196,000	985,000	154,000	770,000
Handling	80,000	240,000	66,000	180,000
Grinding[c]	35,000	100,000	–	–
Calcination[d]	–	–	370,000	900,000
Solids disposal	136,000	400,000	110,000	330,000
Total	447,000	1,725,000	700,000	2,180,000

[a] 3.5% sulfur in coal, limestone at $2.05/ton, and 14.5% capital charges.

[b] Assumes limestone purchased and calcined at power plant.

[c] Grinding cost based on figure 9 in *Sulfur Oxide Removal from Power Plant Stack Gas–Sorption by Limestone or Lime (Dry Process)*, Tennessee Valley Authority, Muscle Shoals, Alabama, p. 37, 1968, with capital charges adjusted from 11% to 14.5% of investment as used in this report.

[d] Calcination cost based on figure 11 in above report; capital charges adjusted from 11% to 14.5% of investment as used in this report.

Table 12. Dust Removal Efficiency, Investment, and Operating Cost of Precipitator-Wet Scrubber Combination in Existing 200-mw Power Unit[a]

Process	System operation	Dust removal efficiency %	Additional oper. cost for system[b] $	Additional new invest. for system[c] $
Basic[d]	Precipitator only	90.0	–	–
A	Two-stage scrubber only	99.5	257,300	1,006,000
A	Precipitator and two-stage scrubber	99.85	293,600	1,075,000
B	Two-stage scrubber only	99.5	260,500	1,030,000
B	Precipitator and two-stage scrubber	99.95	272,500	974,000

[a] Coal burned contains 3.5% sulfur and 12% ash.

[b] Operating cost for scrubber-fan system and for ash-lime transfer system when precipitator is used in process A.

[c] New investment for scrubber-fan system, ash-lime transfer system when precipitator is used in process A, and system for additional solids disposal to pond.

[d] Basic case assumes electrostatic precipitator installed and operating without scrubber; 90% efficiency assumed as noted earlier.

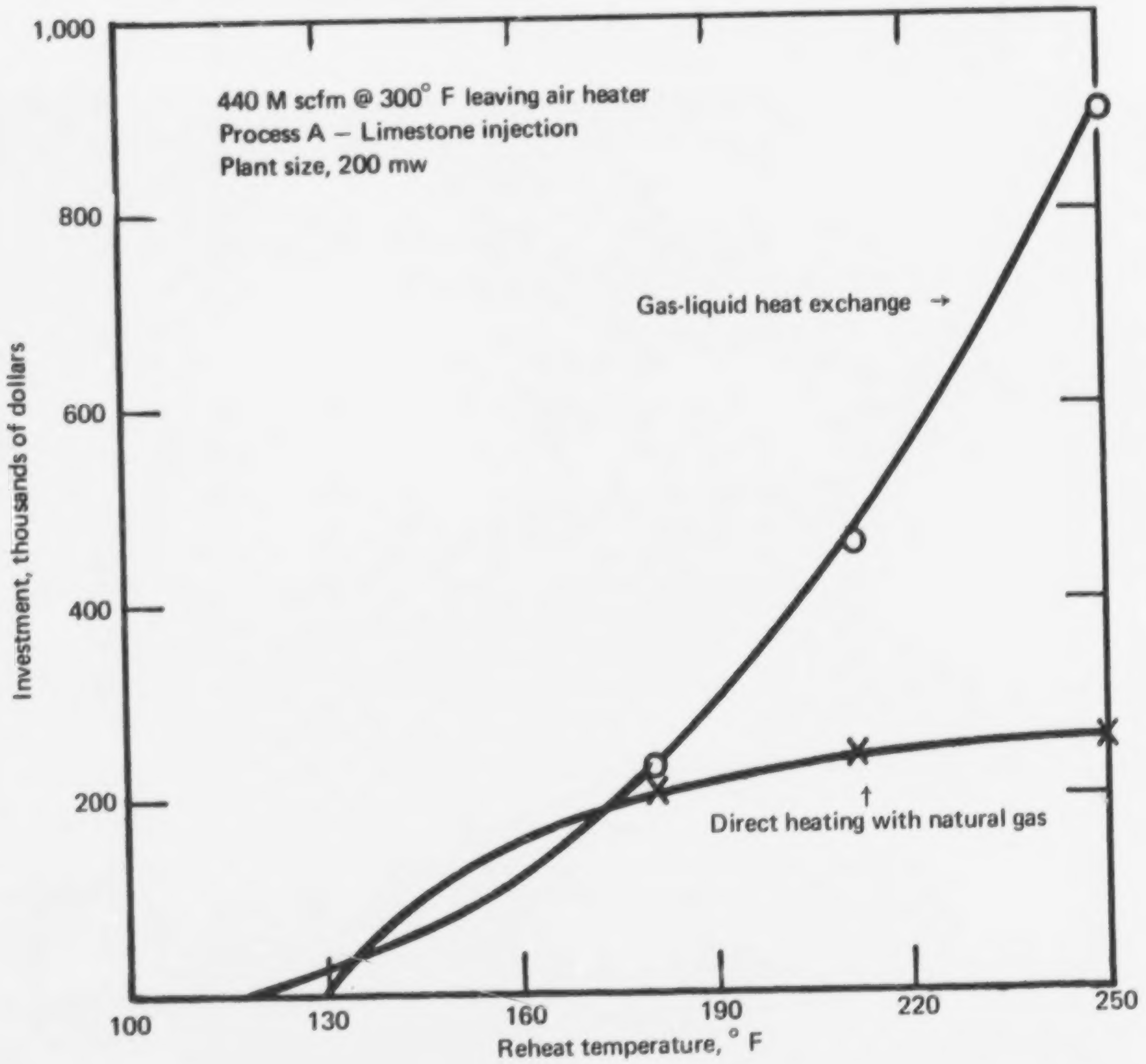

Figure 15. Effect of reheat temperature on investment for stack gas reheating

between the gas streams before and after scrubbing, (4) indirect heating with steam (in a new power plant designed to produce excess steam), and (5) direct heating with coal in a stoker system. In addition, the effects of degree of reheat on operating cost and investment were determined for two of the methods–natural gas heating and indirect liquid-gas heat exchange (Figs. 15 and 16).

A 200-megawatt unit, coal containing 3.5% sulfur, and reheat to 250° F. were assumed for the comparison of reheat methods; the 250°F. was selected arbitrarily because higher levels are quite expensive by the heat exchange methods and do not appear justified from the standpoint of stack gas dispersion. The costs of pressure drop and of quench water to precool at the scrubber were included in the estimate, as these differ with method used. The results (Table 13) show that liquid-gas heat exchange is the most economical and it was selected for the further estimates.

Gas reheat to temperatures lower than 250° F. may be acceptable in some applications of the wet-scrubbing process. Adjustments can be made to the overall investment and operating costs (Figs. 19 and 21), which are based on reheat to 250°F., through use of Figures 15 and 16. For example, for reheat to 190° F. in a 200-megawatt unit by the gas-liquid heat exchange method, the difference in investment (from Fig. 15) between 250° and 190°F. is $600,000 or $3.00 per kilowatt, which should be subtracted from the overall investment; likewise the operating cost (from Fig. 16) would be reduced by $120,000 or $0.20 per ton of coal burned.

If it were necessary to reheat to temperatures higher than 250° F., a combination of reheat methods would be

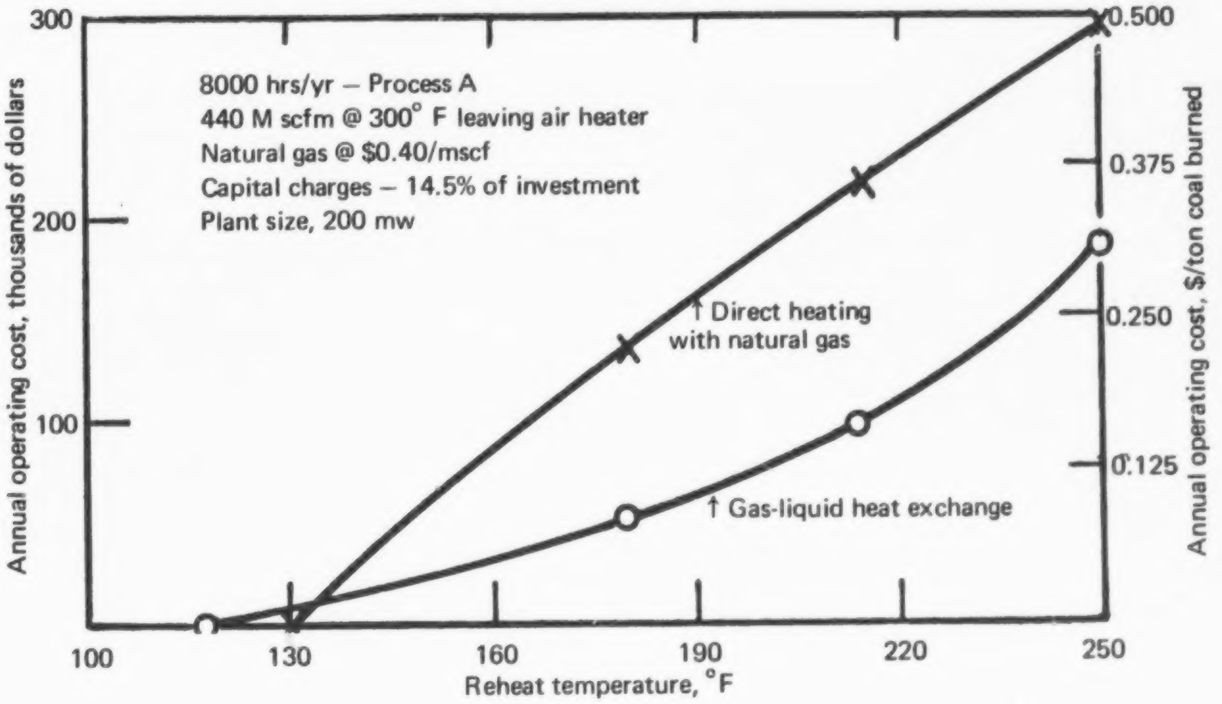

Figure 16. Effect of reheat temperature on operating cost of stack gas reheating

best for economy. The indirect method would be used to reheat up to 230°-240°F. and the natural gas- or coal-fired stoker methods for heating to higher levels. For example, for reheating to 300°F. in a 200-megawatt unit, the operating cost would be $130,000 per year for the indirect portion of the heating and $180,000 per year for direct natural gas for a total of $310,000 per year or $0.52 per ton of coal burned.

In some situations, reheat may be required only when adverse atmospheric conditions reduce stack gas dispersion. For such intermittent operation, the low investment, direct heating methods would be applicable because the cost of fuel, which puts these methods at a disadvantage for continuous use, would be eliminated during nonoperating periods. The effect of reheating operating time (with natural gas) on overall operating costs is shown in Figure 17. Operating cost would be $1.10 per ton of coal with reheat 10% of the time compared with $1.31 for continuous reheat by the cyclic heat exchange method (Fig. 21). However, if the natural gas unit were operated as much as 50% of the time there would be little or no advantage over continuous reheat by heat exchange.

Natural gas was selected for the comparisons in Figures 15 and 16 because it is well proven, flexible, clean, convenient for adding to an existing plant, and low in investment. The coal stoker method is not well proven and reduces the effective sulfur dioxide removal by introducing sulfur dioxide from the coal into the gas. Steam-gas heat exchange is also more economical but is applicable mainly in new power plants because of the steam required.

Sorbent Vendor and Shipping Costs

Vendor Cost: To determine the cost of raw limestone for use as sorbent in processes A and B, a series of f.o.b. price quotations were obtained from vendors located mostly in the eastern United States. The prices quoted are shown in Table 14 and cover varying particle size and calcium oxide and magnesium oxide content. Much of this information was obtained in the previous dry process study and is repeated here for convenience.

It should be noted that these quoted prices are based on costs in existing plants and on an amount of material (50,000 tons/yr.) considerably less than that which would be used by a modern large power plant. If a long-term contract for a very large quantity were entered into, then the vendor might be justified in putting in new equipment and quoting a special low price.

It is not possible, of course, to estimate very closely the purchase price under a long-term contract for a very large quantity of limestone, say, the 386,000 tons per year needed for a 1000-megawatt unit (at 3.5% S in the coal and 110% stoichiometric). A good figure could be obtained only by actual negotiations in a specific situation. However, discussions with limestone producers have supplied a rough

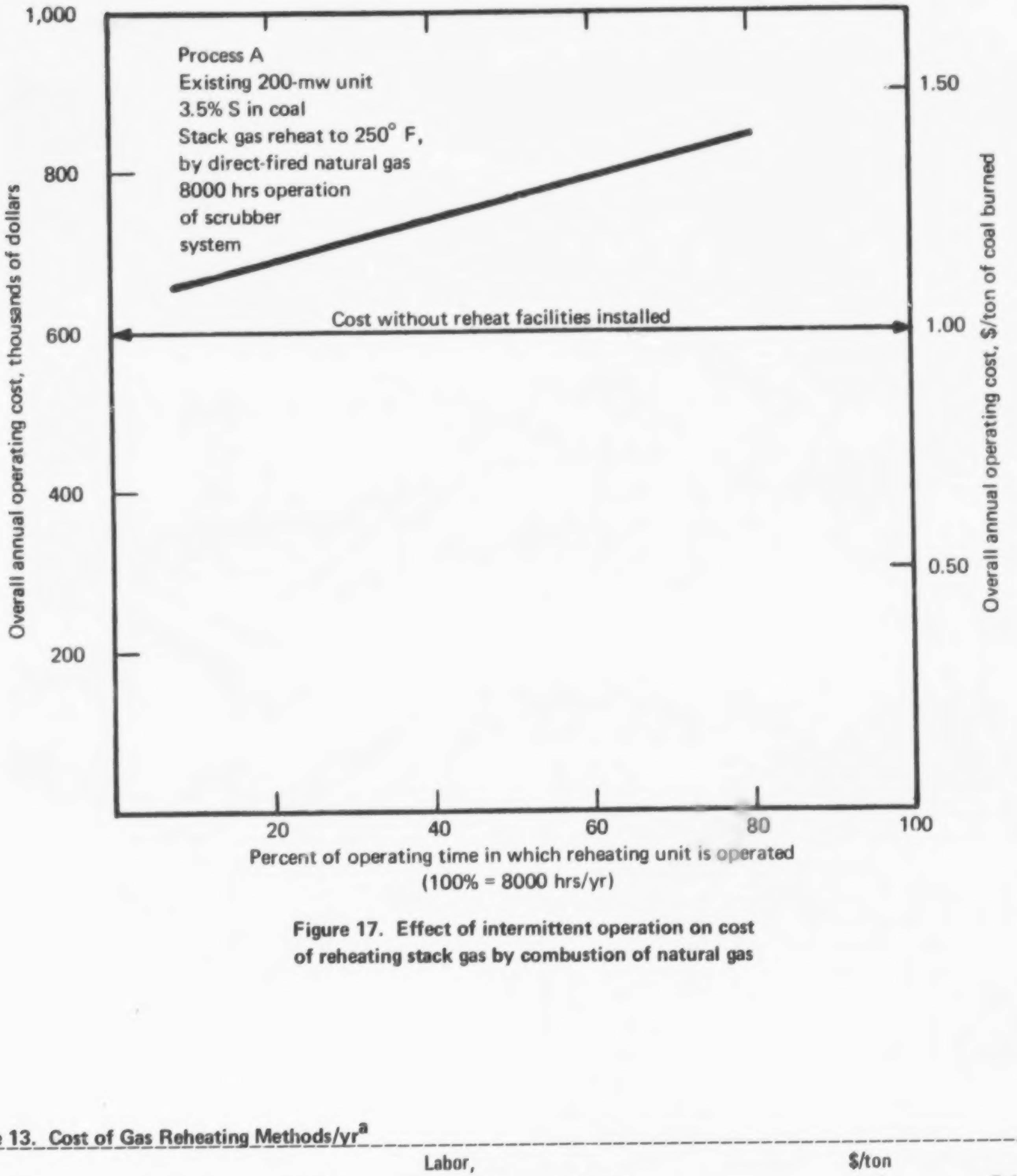

Figure 17. Effect of intermittent operation on cost of reheating stack gas by combustion of natural gas

Table 13. Cost of Gas Reheating Methods/yr[a]

Technique	Fuel, \$	Fan and power, \$	Labor, maintenance & overhead, \$	Capital charges, \$	Total, \$	\$/ton coal burned	Reheat invest., \$
Direct, natural gas	247,000[b]	3,100	5,300	38,400	293,800	0.49	265,000
Indirect, liquid-gas	–[c]	31,000	23,000	131,200	185,200	0.31	905,000
Indirect, gas-gas	–[c]	55,000	64,200	333,000	452,200	0.75	2,300,000
Indirect, steam-gas	147,600[d]	14,000	16,800	60,900	247,300	0.41	420,000
Direct, coal stoker	109,200[e]	3,400	31,000	65,200	208,800	0.35	450,000

[a] 200-mw power plant; reheat to 250° F.; 440 M scfh.

[b] Natural gas at \$0.40/MM Btu.

[c] Indirect liquid-gas and gas-gas utilize waste heat in exhaust gases.

[d] Steam at \$0.30/MM Btu from 200-mw plant designed for excess process steam.

[e] Coal at \$0.20/MM Btu.

Table 14. Quoted Prices of Raw Limestone

	Particle size		Grade, %		F.o.b. $/ton, bulk	
Location	Mesh size indicated	% through given mesh	CaO	MgO	of material	of CaO and MgO
Cherokee, Ala.	¼ in.	100	51.3	1.7	1.35	2.55
Longview, Ala.	200	70	53.5	0.8	3.85	7.10
Ste. Genevieve, Mo.	16	100[a]	55.3	0.4	2.85	5.10
	300	62[b]	55.3	0.4	2.85[c]	5.10
	300	80	55.3	0.4	5.85	10.50
	300	98	55.3	0.4	6.60	11.80
	300	99.6	55.3	0.4	7.60	13.65
Roberta, Ala.	60	50[d]	54.0	–	1.65	3.05
Coleman, Fla.	325	75[e]	54.5	0.6	4.00	7.25
Russellville, Ky.	¼ in.	100	51.7	3.6	2.00	3.60
Durbin, O.	3 in.	100	30.6	21.5	1.50	2.90
	200	48[f]	30.6	21.5	2.50	4.80
Kimballton, Va.	200	82	54.0	2.0	3.00	5.35
Stephens City, Va.	200	75	54.8	0.3	2.75	5.00
Strasburg, Va.	3/8 x 1/8 in.	100	55.0	0.4	2.00	3.60
Adams, Mass.	200	98[g]	54.0	0.7	10.00	18.30
Lucerne Valley, Calif.	200	99	55.0	0.3	7.00	12.60

[a] 21% –80 mesh; sized product.
[b] 100% –35 mesh.
[c] Low price due to product being surplus.
[d] 90% –10 mesh.
[e] 97.8% –100 mesh.
[f] 87% –40 mesh.
[g] 29% –15 microns; 11% –4 microns.

estimate of the potential price. The consensus for a 500,000-ton-per-year supply was about $1.15 per ton; the range was $0.71 to $1.50 per ton.

It is concluded that for relatively small limestone consumption, for example, the 44,000 tons per year needed for a 200-megawatt unit with 2% sulfur coal and 110% stoichiometric injection, or the 36,000 tons (approximately) needed for a 1000-megawatt unit under the same conditions but operated on an intermittent injection basis, the quoted price range of crushed limestone (maximum size, 1/4 in.), $1.35 to $2.00, is a good figure. For larger consumption, however, particularly for the 1000- to 2000-megawatt or larger central stations, much lower cost can be expected. The ultimate might be a situation in which a limestone plant of very large capacity, located adjacent to a navigable river, supplied several large power plants upstream and downstream by barge; this is similar to the supply situation on the Great Lakes, where a large limestone company sells crushed limestone in large quantities at $0.73 per ton f.o.b. ship.

Shipping Cost: Freight costs of limestone by rail or truck transportation are given in Figure 18. Except for localized situations, rail shipment is more economical than truck shipment.

Barging costs are much less than for rail or truck but vary so widely that a curve of cost versus distance would be misleading. Barging is suited mainly to large-volume shipment on a contracted minimum annual tonnage basis, and, of course, to shipping and receiving points on or close to navigable water. Where such a combination exists, very low rates can be obtained; TVA, for example, ships coal in large quantities at an average of about 2 mills per ton-mile. The conditions that make this rate possible may be somewhat favorable, however; the following costs have been estimated as generally applicable to barge shipping of limestone:

	Shipping cost, $/ton		
Tons/yr	100 mi	200 mi	300 mi
150,000	0.50	0.75	0.90
1,000,000	0.45	0.65	0.80

There should be added to the barge rates the cost of unloading at the power plant, since there is little or no unloading cost for trucks of hopper-bottom rail cars. A typical barge unloading cost, based on power plant experience, is $0.16 to $0.17 per ton at an unloading rate of 1000 tons per hour.

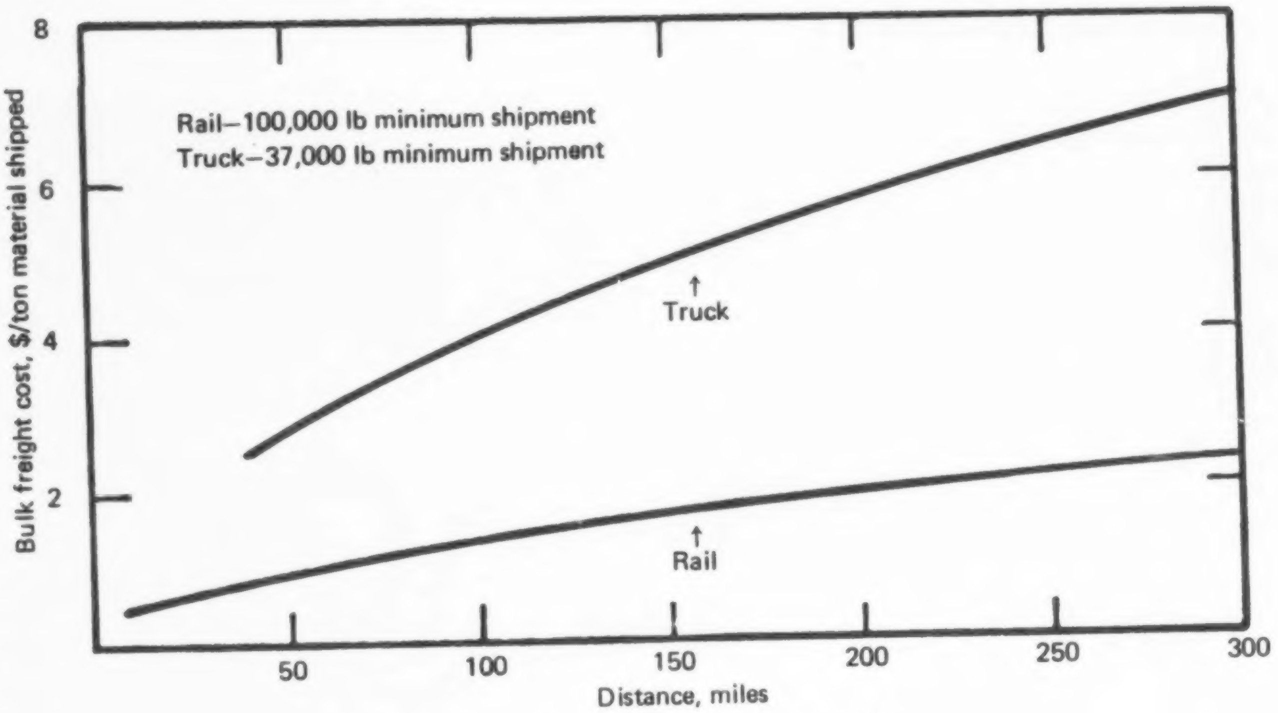

Figure 18. Bulk transportation cost for shipping limestone by truck or rail

Shipping cost can also be reduced by the unit-train method–that is, a special train carrying only limestone and proceeding direct from quarry to power plant. This practice is also generally based on a contracted minimum annual tonnage, and the cost varies widely with conditions. TVA, for example, receives coal by unit-train shipment at rates ranging typically from 4.7 mills per ton-mile for 343-mile shipment to 9.2 mills for 114 miles. Minimum annual tonnage for such contracts is 1 to 2 million tons; minimum trainload tonnage is about 5000 tons.

In comparison with Figure 18, which is based on single carloads, the shipping cost for the unit train is about 30% lower.

As previously mentioned, a Colbert County, Alabama, location has been assumed for the purpose of evaluating the economics of limestone–wet scrubbing. Based on a quotation from a vendor located in the Colbert County area, the assumed f.o.b. cost of 1/4- by 0-inch limestone used in the overall process evaluations presented herein is $1.35 per ton. Trucking cost to the power plant would be $0.70 per ton, giving a delivered cost of $2.05 per ton of raw limestone.

Overall Process Investment Evaluation

The overall fixed investments for processes A and B in existing plants of various sizes (3.5% S in coal) are summarized in Table 15 and shown graphically in Figure 19. In addition, curves in Figure 19 show the investment as dollars per kilowatt of power generation capacity, a common power industry method of expressing investment.

Table 15. Overall Investment for Scrubbing Facility in Existing Plant

	Unit size, mw					
	200		500		1,000	
Process	$	$/kw	$	$/kw	$	$/kw
A	2,610,000	13.05	5,425,000	10.85	8,210,000	8.21
B	2,760,000	13.80	5,830,000	11.66	8,820,000	8.82

Investment for process A includes equipment for limestone receiving, storage, conveying, dry pulverizing, injection into the boiler; two-stage gas scrubbing; indirect liquid-gas cycle reheat of scrubbed flue gas to 250°F.; direct pumping of 10% waste solids to pond with no return of water; and incremental pond storage of the product solids. Investment for process B includes the same equipment as for process A except that wet grinding and slurry preparation replace dry pulverizing and injection into the boiler.

The overall investments for processes A and B installed in new 1000-megawatt plants are shown below. Also given is the effective investment after credit is taken for eliminating the electrostatic precipitator (99% efficiency

	Overall investment for scrubbing facility in new plant (1,000 mw)			
	Actual		After precipitator credit	
Process	$	$/kw	$	$/kw
A	7,620,000	7.62	6,320,000	6.32
B	8,270,000	8.27	6,970,000	6.97

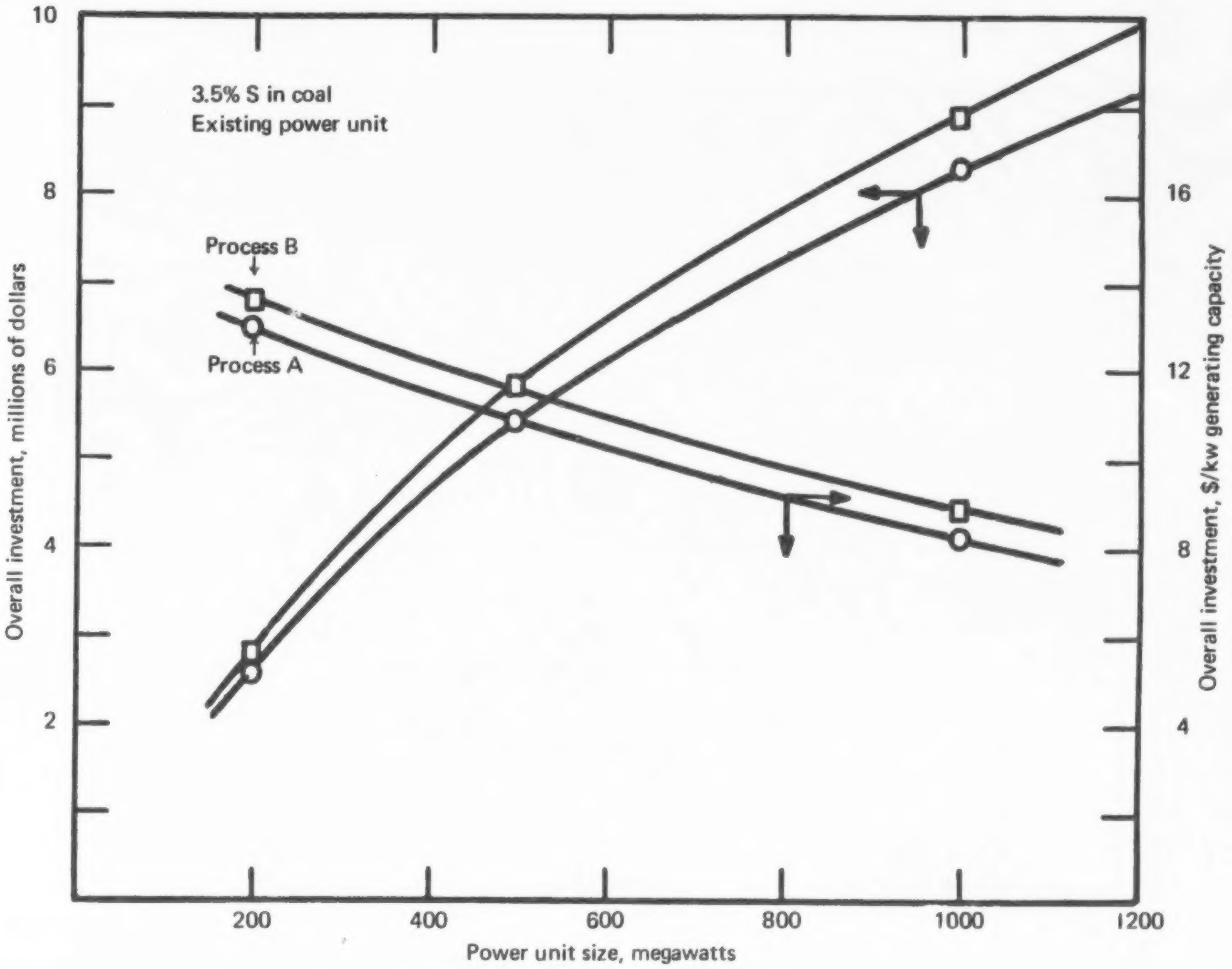

Figure 19. Effect of power unit size on limestone—wet scrubbing investment

assumed). This estimate shows the double advantage of using the wet-scrubbing method in new rather than old power plants. Actual investment is about 9% lower in a new 1000 megawatt plant and the credit for eliminating the precipitator gives a further reduction of 14%, or 23% overall.

In all cases, the investment required for process A is less than that for process B. Major differences are in raw material and waste handling due to the different raw material requirements of the two processes (process A, 110% stoichiometric; process B, 130% stoichiometric) and in the requirement that a portion of the heat transfer surface for process B stack gas reheat be of alloy steel for corrosion resistance, whereas process A requires only regular materials.

The effect on investment of sulfur content of coal burned, in a 200-megawatt existing unit equipped with process A, is shown in Figure 20. The investment is only slightly sensitive to sulfur content because the gain is limited mainly to size economy of raw material and waste-handling systems. Doubling the sulfur content in the coal (2-4%) increases the total investment only 20%.

Overall Operating Cost Evaluation

A summary of the overall operating costs of processes A and B at various unit sizes (existing plants and 3.5% S in coal) is presented in Table 16 and shown as curves in Figure 21. The data relate both total annual cost and cost per ton of coal burned to the unit size.

The overall operating costs for processes A and B used in a new 1000-megawatt power unit are given below. These values reflect the capital charge credit for precipitator

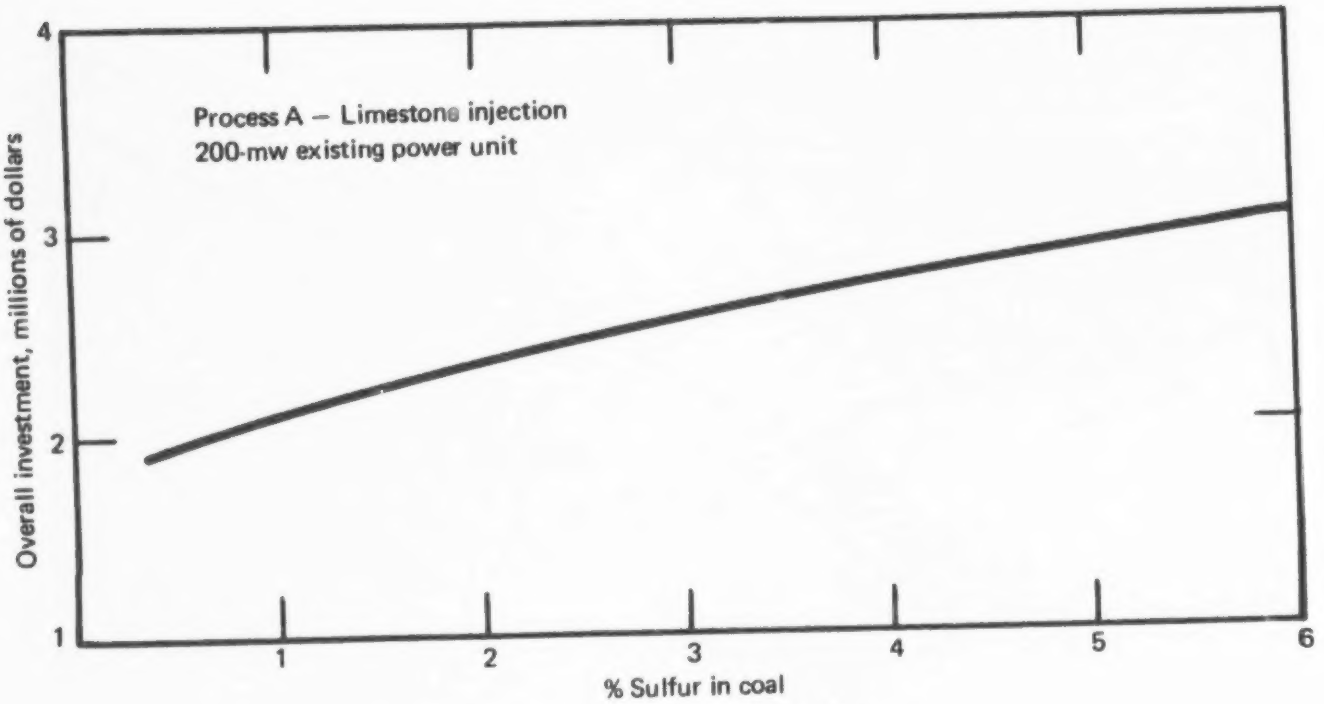

Figure 20. Effect of sulfur content of coal on limestone–wet scrubbing investment

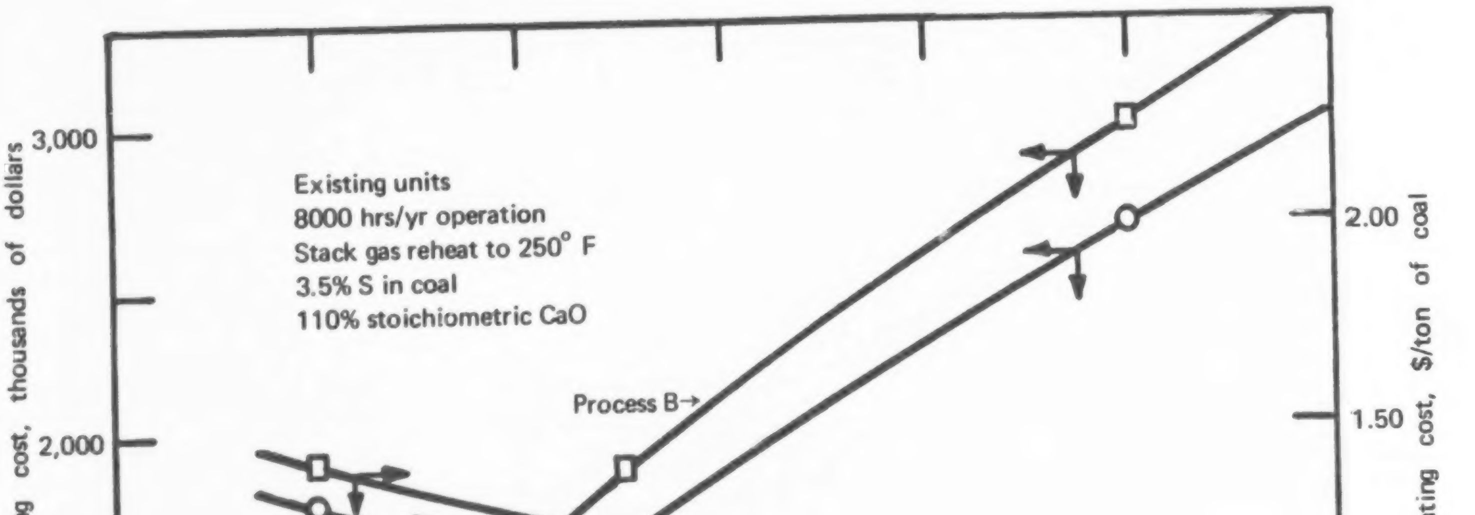

Figure 21. Effect of power unit size on limestone–wet scrubbing operating cost

investment and the economy of a large, new installation as previously described.

	Overall operating cost (1,000-mw new power unit)	
Process	Total annual cost, $	Cost, $/ton of coal burned
A	2,327,100	0.775
B	2,601,000	0.866

The effect on operating costs of sulfur content in the coal burned, in existing 200-megawatt units using process A, is shown in Figure 22. Doubling the sulfur content of the coal (for example from 2-4%) increases the overall cost by 37%.

Operating cost sheets detailing the above estimate summaries are given in Appendix C.

The method of deriving the assignable credit for electrostatic precipitator operation is described in Appendix C. Also included in Appendix C is a discussion of the credit for corrosion reduction when limestone is injected into a boiler as in process A. The method for calculating thermal loss due to limestone injection is described in the dry process report.

To complete the operating cost evaluation, the effects of raw material cost and annual scrubber operating time are shown in Figures 23 and 24. Figure 24 applies to intermittent operation of the scrubber system only and continuous operation (8000 hr./yr.) is assumed for the boiler. The effect of boiler load factor is also important; here the cost of sulfur removal per unit of power produced increases with decrease in operating time—in contrast to the saving achieved by intermittent operation of the sulfur removal system when the boiler is operating full time. In both cases, high investment is a handicap because the capital charges continue when the scrubber system is out of operation. The effect of boiler load factor is shown in Figure 25.

Table 16. Overall Operating Cost (Existing Units)

	Unit size, mw					
	200		500		1,000	
Process	Total annual cost $	Cost $/ton coal burned	Total annual cost $	Cost $/ton coal burned	Total annual cost $	Cost $/ton coal burned
A	784,100	1.308	1,671,800	1.114	2,717,400	0.901
B	858,800	1.421	1,862,300	1.241	2,993,000	0.986

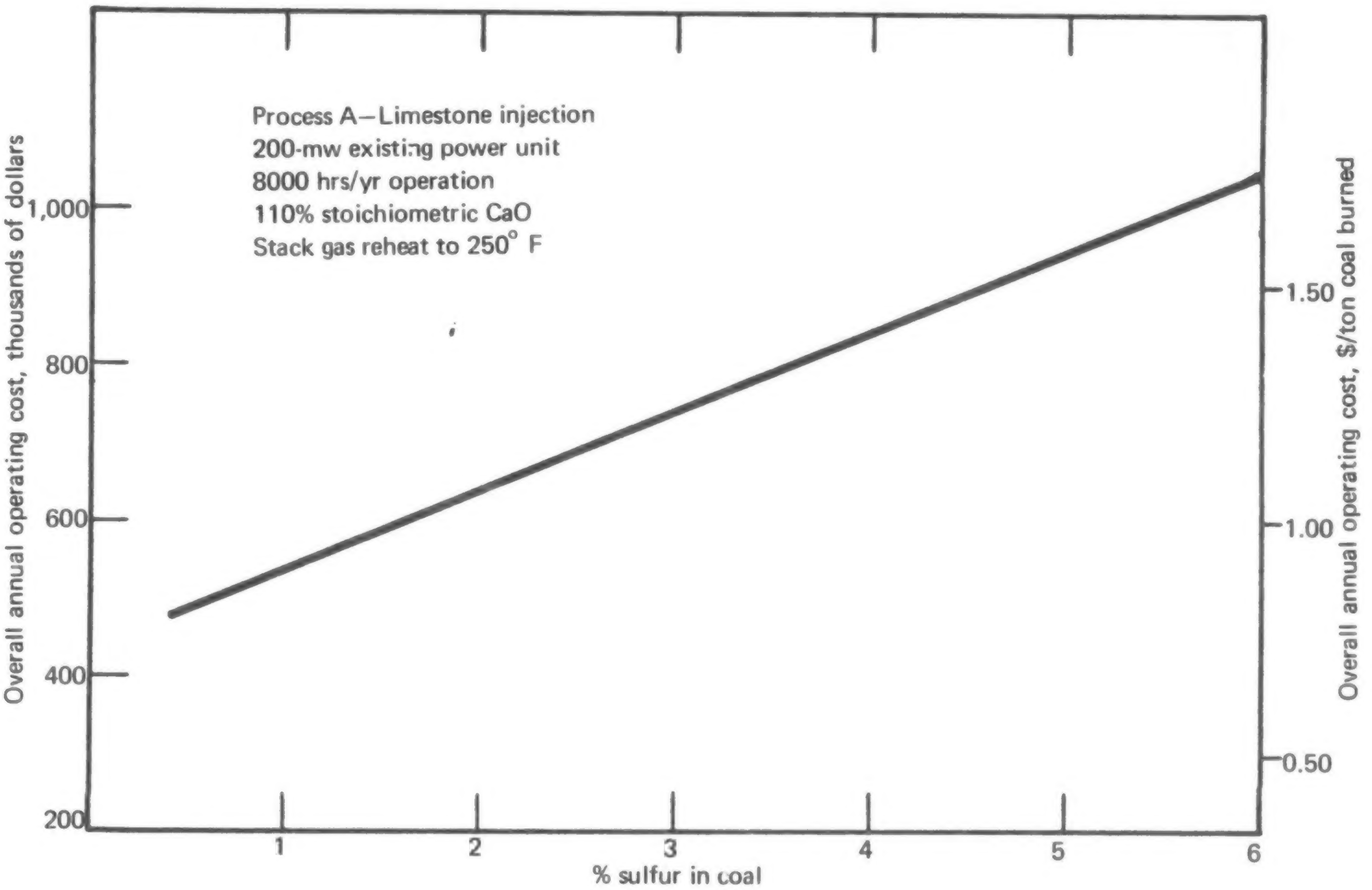

Figure 22. Effect of sulfur content of coal on limestone—wet scrubbing operating cost

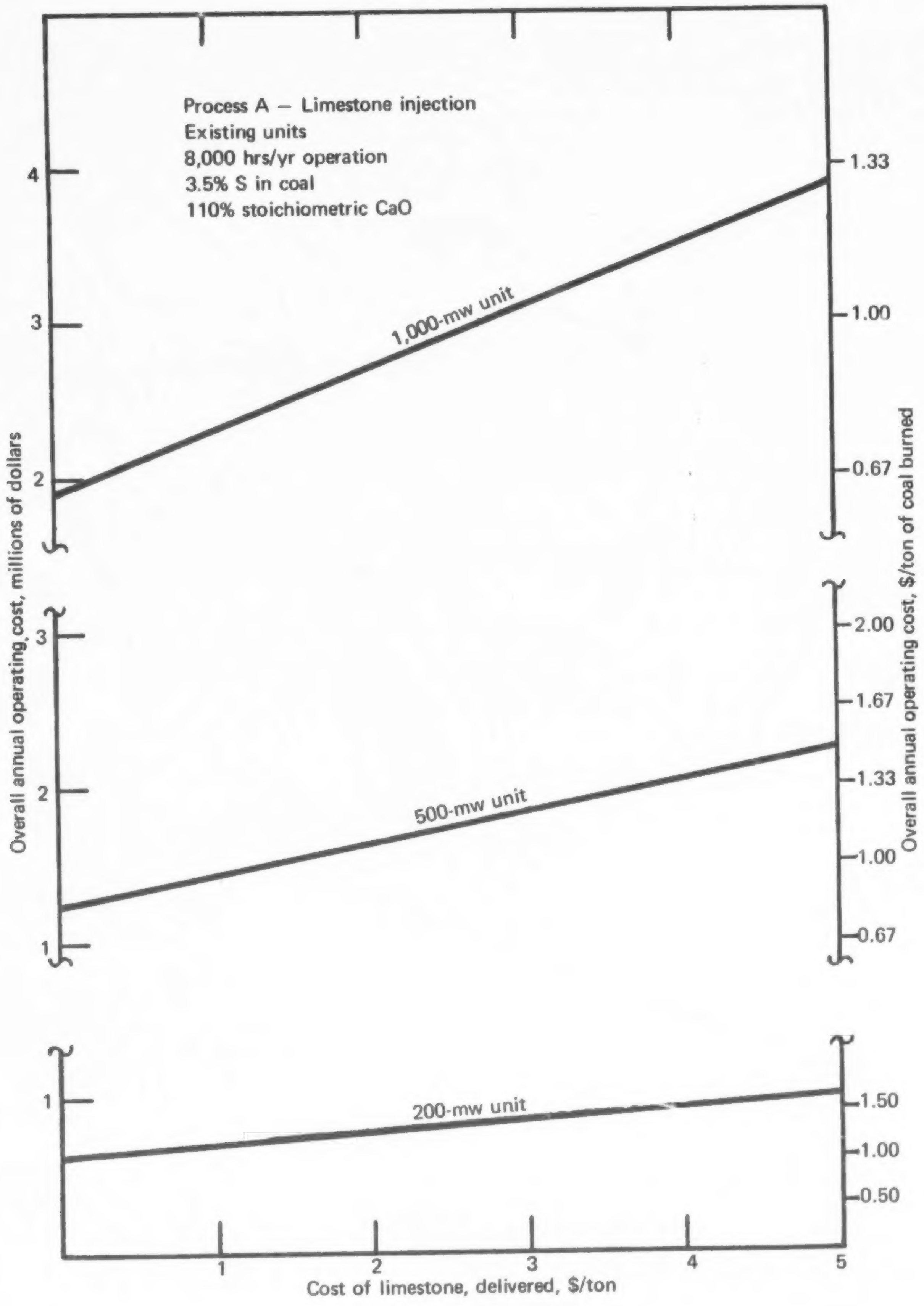

Figure 23. Effect of limestone cost on limestone–wet scrubbing operating cost

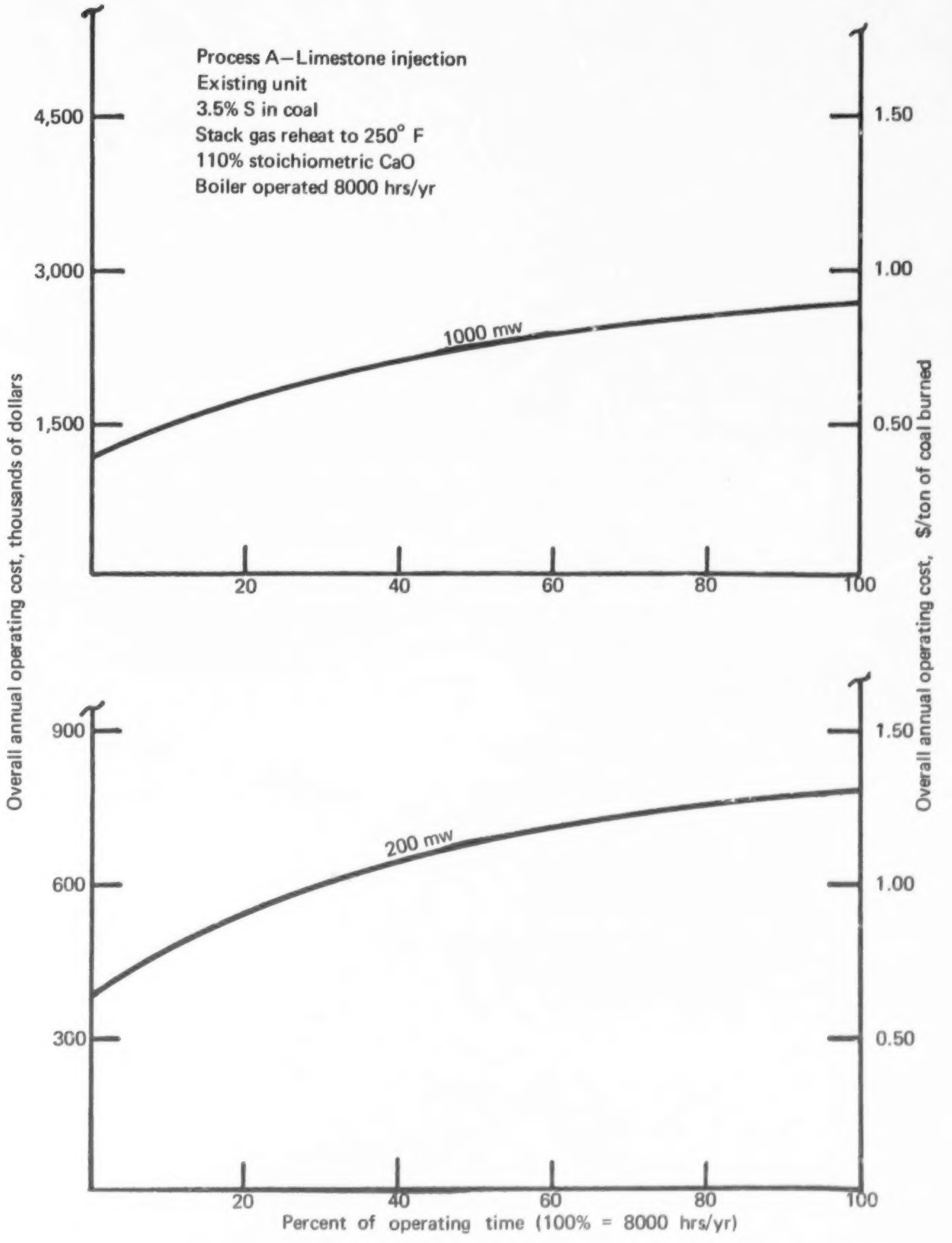

Figure 24. Effect of scrubber system operating time on limestone—wet scrubbing operating cost

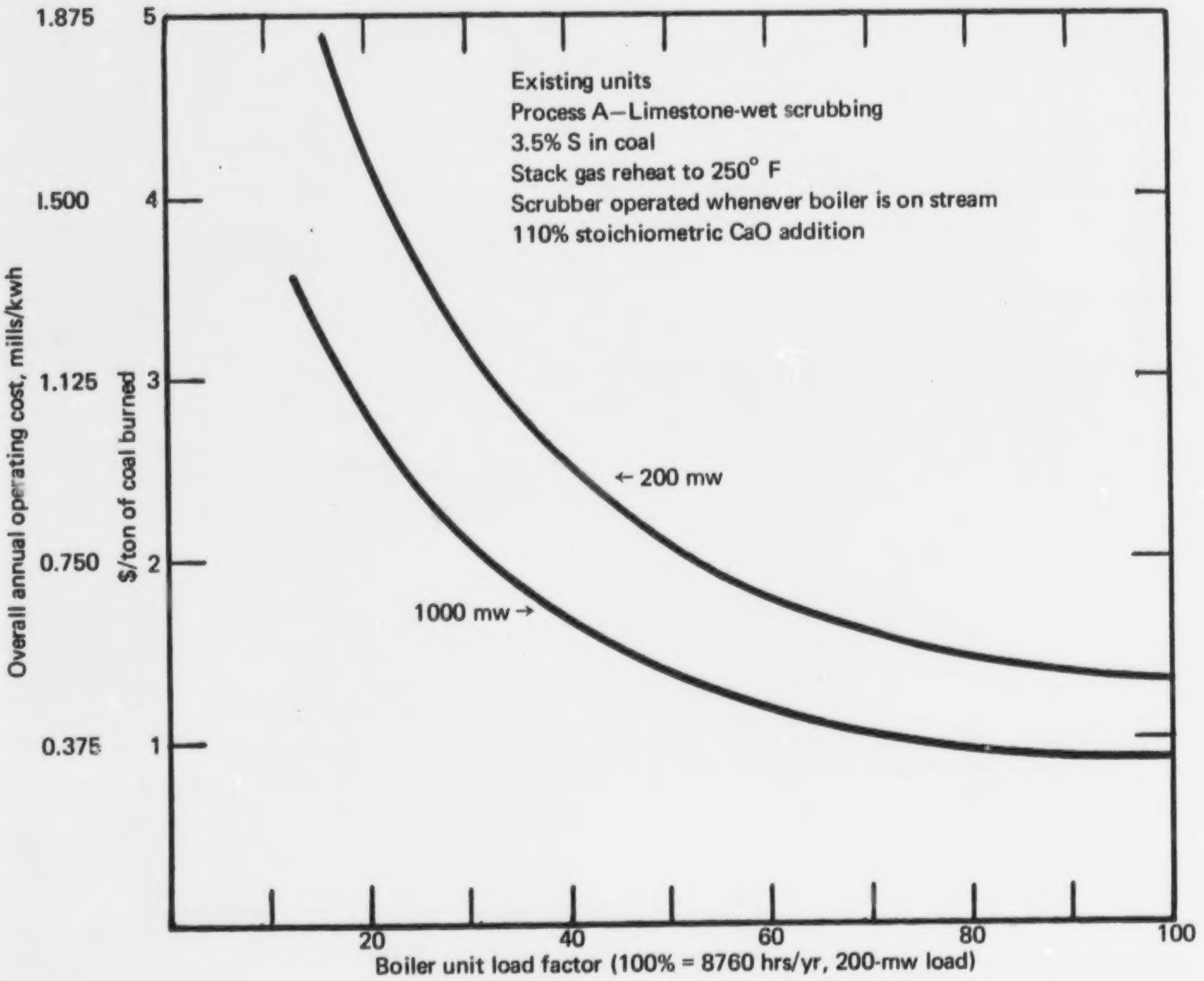

Figure 25. Effect of boiler load factor on limestone—wet scrubbing operating cost

Research and Development Needed

Many of the assumptions for the design and cost studies were made with limited information. In order to improve the accuracy of estimates, further information is needed in the following areas.

Absorbent Efficiency

Further data are needed on the several parameters involving properties and degree of utilization of the absorbent:

1. Calcitic versus dolomitic limestone.
2. Effect of overheating limestone injected into the boiler.
3. Particle size.
4. Variations between limestone types.
5. Oxidation of sulfite to sulfate.
6. Dissolution rate of absorbent, particularly of limestone introduced directly into scrubber recirculating liquor.

Some of these data will be generated by the full-scale wet-scrubbing systems being installed commercially. Also, a pilot plant wet limestone scrubbing program to be conducted under a NAPCA contract should produce valuable data on absorption efficiency.

Research is needed particularly on the problems of using the limestone–scrubber addition process. If good absorbent efficiency could be obtained, the process could well be superior to the injection-scrubbing method–in better boiler operation, lower operating cost, and avoidance of scaling. Possible approaches to increasing the rate of dissolution, which presumably is the limiting factor, are as follows:

1. Grinding to small particle size.
2. Increasing the delay time between mixing of the limestone feed with scrubber effluent liquor and introduction of the resulting slurry into the scrubber.
3. Introducing carbon dioxide (as stack gas) into the slurry during or after limestone addition.
4. Prescrubbing with the makeup water to give an acidic solution for dissolving the limestone.

The problem of absorbent efficiency, as well as all the other problems of the limestone–wet scrubbing method, would be more amenable to solution if more were known about the chemistry involved. Research is urgently needed on both reaction kinetics and equilibrium compositions. The latter is currently under study at TVA, in a computer program in which known equilibrium constants are being used to determine equilibrium concentrations in the five-component system $CaO\text{-}CO_2\text{-}SO_2\text{-}SO_3\text{-}H_2O$. Further work of this type is needed, especially with MgO added, and the mechanisms of transfer of reactants into the liquid phase should be established.

Scaling

Although a satisfactory method for preventing scaling was developed by Howden-ICI, scaling and plugging is a major problem in the current tests by Union Electric and Combustion Engineering. Research and development should proceed along the following lines:

1. Determine if introduction of the absorbent directly into the scrubber is a major cause of scaling.
2. If so, test arrangements for separating the lime from the gas ahead of the scrubber and introducing it into the scrubber liquor circuit after the scrubber. For an existing plant, the existing precipitator could be used to separate the lime. For a new plant, two-stage scrubbing with a venturi used to remove the lime in the first stage should be tested.
3. Develop further data on the Howden-ICI process (introduction of absorbent into the recirculating liquor) to determine if it is adaptable to the more efficient modern types of scrubbers and to various types of limestone.
4. Study further the chemistry of the system in respect to equilibrium composition, degree of supersaturation, rate of desupersaturation, and adherence of precipitated salts to various types of scrubber surfaces.
5. Test dolomitic limestones to determine if the scaling problem is affected by the presence of magnesium salts.

Corrosion

The effect of limestone injection on high-temperature corrosion in the boiler will be studied in full-scale tests of the dry limestone process (discussed in the dry process report). However, low-temperature corrosion may occur in the heat exchanger installed before the scrubber if the limestone is added in the scrubber liquor circuit. It is assumed that corrosion will be no worse than in operation of a low-level economizer (22) and that use of Corten alloy will be acceptable. Data to confirm this are needed.

Control of scrubber solution pH is necessary for corrosion protection of the circulating system. Use of corrosion-resistant lining will reduce the hazard of damage to the scrubbers but pH variation could cause corrosion to pumps and piping that would not be seriously attacked by the normal scrubber solution at a pH of about 6.2. Plant operating data are needed to evaluate this potential problem.

Oxidation

On the basis of experience reported, approximately half of the sulfur dioxide removed in the scrubber is oxidized to sulfate. The remainder forms calcium sulfite which could, under some conditions, cause an unacceptable depletion of dissolved oxygen in receiving streams. A study of methods for promoting oxidation in the scrubber is needed, plus determination of the effect of the more complete oxidation on scrubber performance.

A consideration beyond the scope of the present study is the possibility of regenerating the reaction products to reduce limestone requirement and to recover the sulfur value of the sulfur dioxide. NAPCA is sponsoring work in this area. Regeneration of calcium sulfite should be more practical than decomposition of calcium sulfate because the energy requirement is lower. If regeneration is an objective, a scrubbing process that would inhibit oxidation would be desirable. Use of oxidation retardants might have merit, particularly for the process in which limestone is added in the scrubber circuit and calcium sulfate formation in the boiler therefore is avoided.

Waste Disposal

It has been assumed that the general practice of sluicing fly ash to settling ponds and overflowing clarified effluent to a receiving stream could be used for limestone wet scrubber effluent if desired. Further information is needed on the composition and other properties of the slurry to evaluate this assumption. For example, the sulfite-sulfate distribution in the reaction products will have a major effect on the content of dissolved solids, as the sulfates of calcium and magnesium are about 50 and 70 times as soluble, respectively, as the sulfites. Moreover, the particle size distribution of the suspended solids is not well defined and therefore settling rates cannot be accurately predicted. Abrasive characteristics of the fly ash-lime mixture are also uncertain. Finally, the feasibility of recycling the pond water, which will likely be desirable in most cases and essential in some, should be explored.

Intermittent Operation

The limestone–wet scrubbing method would be greatly improved if it could be made more amenable to intermittent operation. The best opportunity for this is in an existing plant, where the existing precipitator could be operated in series with the scrubber to remove dust so that liquor flow to the scrubber could be shut off when sulfur oxide recovery was not required. The feasibility of operating the scrubber in this way should be explored.

For a new plant, where the scrubber would be designed to remove both dust and sulfur oxides, liquor flow could not be cut off. However, the lime feed could be stopped and the dust scrubbed out with the resulting sulfurous acid. The feasibility of this should be studied.

Nitrogen Oxide Removal

Although the main objective of limestone–wet scrubbing is sulfur dioxide removal, dust is also removed and if nitrogen oxide absorption could be improved enough to give adequate removal the wet process would virtually solve the power plant emission problem. Present information indicates that about 20% absorption of nitrogen oxides is all that can be expected with the current process. It might be possible to improve absorption by identifying the pertinent operating conditions and adjusting them to promote nitrogen oxide removal, or by adding another scrubbing stage in which an absorbent with higher affinity for nitrogen oxides could be used.

Work in this area would be a logical extension of on-going or planned research and development efforts on use of limestone in a wet-scrubbing process.

Partial Removal of Sulfur Oxides

One of the major questions in regard to the limestone–wet scrubbing process is how it compares economically with the dry injection method. It is difficult to compare the two, however, because the wet process is quite efficient, capable of removing 95% or more of the sulfur oxides, whereas the dry method probably cannot do better than 50 to 60%.

Actual cost of operation obviously is not a good basis for comparison because it does not take into account the higher efficiency of the wet process. A more logical approach is to design or operate the wet-process unit for only 50 to 60% removal and compare the resulting costs.

Comparison on this basis is complicated. It should be noted in the first place that unless full reheat is assumed the wet process must operate at higher than the 50% efficiency assumed for the dry process to give the same ground concentration. For reheat to 250° F., removal by the wet process would have to be about 60%. However, the technical feasibility of only 60% removal may be questionable. A scrubber designed for low sulfur dioxide removal would not be a good dust collector and reducing the limestone addition to an efficient scrubber could cause acidic conditions that would likely result in both poor absorption and corrosion. Use of a low-efficiency scrubber in a plant equipped with an electrostatic precipitator might be practical, but scrubber design and operation could well be difficult. A delicate balance would have to be maintained to ensure the proper degree of reaction between sulfur dioxide and lime; if much of the excess sulfur dioxide reacted, the system would become acidic.

In a multiplant station, use of the wet-scrubbing process might permit installation of gas-cleaning facilities on only half of the units as a means of matching the assumed 50% removal for the dry method installed on all the units. For example, if only one unit of a station made up of two 200-megawatt units were equipped with wet scrubbing, the overall removal (assuming 95% removal in the wet scrubber) would be 47.5% and cost per ton of coal would be reduced by 40% or so as compared with the dry system. However, this would require good mixing of the untreated and treated plumes before they reached the ground, which would not be assured for all atmospheric conditions. Moreover, if the plumes did mix, the resulting single plume should behave in accordance with the curves in Figure 3, and the temperature drop induced by the treated plume would have an adverse effect on the degree to which ground-level concentration is reduced. Assuming an average combined plume temperature of 275° F., ground concentration would be reduced only by about 38%.

Hence the assumed 50% sulfur dioxide removal by the dry process, and the accompanying 50% reduction in ground concentration, could not be duplicated in a wet scrubbing installation merely by scrubbing the gas from alternate boiler units. The difficulty could be avoided by passing about 15% of the gas from the unequipped boiler to the scrubbing system on the other boiler and sizing the scrubber accordingly. This would bring in other problems, and trying to evaluate them would appear to be carrying the string of conjecture too far.

Another approach (for an existing plant) is to divide the gas from the air heater into two streams, one (about 60% of the total gas) passing through the scrubber and the other through the existing precipitator. The two streams would then be recombined and passed into the stack.[9] The estimated operating cost of such a system is $0.82 per ton of coal (200-mw. unit) as compared with $1.05 for the dry process.

While this is a promising potential reduction in cost, the arrangement would require that the ash-lime mixture from the precipitator be introduced into the recirculating scrubber slurry (to use the unreacted lime), which could introduce operating problems because all the fly ash would then go into a scrubber much smaller than for the normal design. Moreover, for a new plant the advantage of eliminating the precipitator could not be realized.

The problem would be simpler if process B (limestone introduced into the scrubber circuit) were used, as the ash from the precipitator could then be sent directly to the pond. However, process B is more expensive than injection-scrubbing and would require that more gas be passed through the scrubber, the latter because sulfur dioxide removal in the boiler would not be obtained and therefore the concentration in the gas stream diverted through the precipitator would be higher.

It is concluded that partial removal of sulfur oxides by wet scrubbing probably is feasible but that potential problems would have to be worked out. If partial removal were acceptable as the basis for design, a power producer could put a scrubbing unit in parallel with an existing precipitator to give the degree of sulfur oxide removal required by pollution authorities. Successful development of such an arrangement would make wet scrubbing more economical than the dry process, even for small plants, and therefore development work should be carried out. Pilot plant testing may be feasible, but tests in an operating unit would be preferable.

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[9] The precipitator and scrubber could not be operated in series, with the gas stream split before the scrubber, because the overloaded precipitator would not remove all the dust and therefore the stream bypassing the scrubber would carry dust out the stack.

Conclusions and Recommendations

The main advantages of the limestone–wet scrubbing process are (1) capability of highly efficient removal of both sulfur dioxide and dust, (2) low investment as compared with processes that recover sulfur products, and (3) since a throwaway product is produced, no involvement of the power company in marketing problems. Disadvantages are (1) cooling of the plume, (2) potential contribution to water pollution problems, (3) need for solids disposal, and (4) no return to offset increased operating cost. Compared with the dry limestone process, wet scrubbing is more efficient and incurs lower waste disposal costs because of more efficient utilization of limestone. However, investment is higher and the process is less adaptable to intermittent operation, particularly for new plants in which the scrubber would be the only dust collector installed.

Investment and operating costs are slightly higher when limestone is added directly into the scrubber circuit, mainly because of the higher limestone requirement assumed and the higher cost of the corrosion-resisting reheating equipment required. Use of lime would reduce the absorbent requirement but the cost of calcining limestone to lime outside the boiler is prohibitive. The advantage of introducing the absorbent into the scrubber circuit is that potential boiler operating problems inherent in the limestone injection method are avoided, scaling is easier to control, and operating cost may be lower.

The adverse effect of plume cooling is difficult to evaluate. A study of the effect of stack temperature on plume disperson shows that a high reduction in ground-level sulfur oxide concentration is obtained, even with a cooled plume, if the scrubber operates efficiently and removes most of the sulfur oxides (say, 80% or more) from the gas. However, as the removal of sulfur oxides (or any other pollutant) is decreased the reduction in ground concentration drops rapidly and at about 45% removal the ground concentration is not reduced at all (for a plume at scrubber exit temperature); in other words, the plume cooling completely nullifies pollutant removal in the scrubber. This is significant mainly in regard to nitrogen oxides, which are not removed to any great extent in the scrubber. If the plume is not reheated, the scrubbing would, in effect, increase the ground concentration of nitrogen oxides.

A further consideration is that adverse effect of plume cooling will quite likely be increased if there is any mist present in the gas as it leaves the stack; mist can be present because of condensation between the scrubber and stack top and because of incomplete removal of scrubber spray in the mist eliminator.

It is recommended, therefore, that the stack gas be reheated at least enough to ensure absence of mist in the gas leaving the stack. Whether or not reheating to a higher temperature is justified, and what temperature level should be selected, will depend on the particular pollution situation involved. The high cost of reheating to higher than about 250°F., however, would make temperatures above this level difficult to justify.

A comparison between the wet and dry limestone processes is difficult to make because of the lower efficiency of the dry process, assumed to be no higher than 50 to 60%. In situations where a high degree of sulfur oxide removal, say over 80%, is essential, only the wet method would be suitable. In contrast, there may be many situations in which the lower degree of removal is adequate and either process could be used. In comparing the economics of the two, however, it is difficult to quantify the advantage of the higher degree of sulfur oxide removal in the wet process. Hence the economics will first be compared without attaching any value to degree of absorption.

For continuous operation, wet scrubbing appears to be a better choice than the dry process for the larger plants. For a 1000-megawatt existing plant operating 8000 hours per year and burning coal containing 3.5% sulfur, investment for the dry process (200% stoichiometric limestone) would be $3,950,000 or $3.95 per kilowatt and the operating cost would be $2,747,900 per year or $0.91 per ton of coal burned. Use of the wet-scrubbing system in a similar situation (with 110% of the stoichiometric amount of limestone injected into the boiler and the stack gas reheated to 250° F.) would more than double the investment ($8.21/kw.) but the total operating cost for continuous operation would be $0.90 per ton of coal. Since capital charges are included in the operating cost, the higher investment is not a handicap unless the unit is operated on an intermittent basis.

Hence wet scrubber economics, considering the other advantages of the method, are favorable for a large plant; for a new plant they would be even better because credit could be taken for eliminating precipitator investment. The disadvantages– lower stack temperature and possibly more water pollution–are difficult to quantify. If air quality considerations dictated full reheat to normal stack temperature (about 300° F.), the overall cost would be increased to $1.11 per ton of coal. And if it developed that use of dolomite made pond water recycling mandatory because of stream pollution, the cost would be increased in areas where dolomite is plentiful but high-calcium limestone is not.

On the other hand, the cost of wet scrubbing could be much lower than the $0.90 per ton of coal given above if lower stack temperature were acceptable. If it is assumed that the minimum acceptable temperature is that required to avoid the presence of water mist in the gas leaving the stack, and that this temperature is, say, 175° F. (the exact temperature varies with plant design and operating conditions), then the wet-scrubbing cost for the 1000-megawatt unit would be only $0.78 per ton of coal. And if a shorter stack should be acceptable to pollution control authorities, the resulting credit would reduce the overall cost even further.

These considerations show the wide range of costs that might be involved in applying the wet-scrubbing process to various plants, which makes economic generalization difficult. In contrast, the dry process is much less affected by cost variables.

The wet process would gain in the comparison if costs were reduced by operating at lower efficiency, on the basis that if, say, 50% removal by the dry process were acceptable in a particular situation, then comparable efficiency should be adequate for the wet process. The problems of partial removal were discussed in the research and development section. Several plant arrangements can be considered but each introduces problems. The best approach for an existing plant appears to be operation of the precipitator in parallel with a scrubber designed to handle about 60% of the gas by the injection scrubbing method; the only apparent problem is that all the ash would end up in the recirculating liquor of the relatively small scrubber. This may not be a serious problem, and the savings—$0.82 per ton of coal against $1.05 for the dry process—are considerable.

The dry system has the advantages of simpler operation, less tendency to operational upset, and having operating steps that are similar to those already practiced in the power plant (grinding, injection, dust removal). The wet process is a somewhat complicated chemical operation, requiring closely controlled operation of a type foreign to power plant practice. However, these differences are not subject to quantitative evaluation.

If intermittent operation were feasible as a means of reducing cost, the dry system should be superior (assuming that 50 to 60% removal is acceptable). The dry process cost for intermittent operation (total of 30 days/yr. in 200-mw. unit) is $0.47 per ton of coal for all coal burned per year; see dry process report). As noted earlier, the mechanics of intermittent operation would present some problems in the wet-scrubbing method. Moreover, even if technically feasible, cost reduction for the 30-day operation would reduce total operating cost only to $0.75.

The same reasoning applies to older plants (usually of small size) that operate intermittently as peak-load units. The effect of boiler load factor on cost of sulfur oxide removal for the limestone—wet scrubbing process as compared with the dry process is shown in Figure 26. Because of the lower investment required, the dry process is not penalized as much by low boiler load factor as is the wet method. For example, the dry process increases cost of power by about 0.41 mill per kilowatt hour at 100% boiler load factor and by 0.56 mill at 50% load factor, an increase of 0.15 mill between the two levels. In contrast, the increase in cost from 100 to 50% load factor for the wet process is about 0.28 mill.

It is concluded that both the wet and dry processes have their place, the wet method particularly for large, new plants and the dry technique for small, peak-load plants or for any plant so located that intermittent operation is acceptable. Hence the wet process becomes the basic method for comparing with recovery processes, which also are more appropriate for large baseload plants than for units operated intermittently. The costs presented in this report indicate that because of the high and increasing cost of low-sulfur fuel, limestone—wet scrubbing will generally be the most economical method for sulfur oxide control—short of entering into the uncertainties and complications of recovery. In other words, for large plants the power producer likely will adopt wet scrubbing unless a recovery process shows promise of being more economical, that is, a net cost sufficiently lower to make the gamble inherent in recovery processes acceptable.

The present study suggests that the forthcoming pilot plant program should include study of the following:

1. Variation of absorption efficiency with limestone type and with calcination conditions in the boiler.
2. Variation of absorption efficiency with scrubber type.
3. Effect of particle size on absorbent utilization. This is of particular interest for direct addition of limestone in the scrubber circuit.
4. Use of the delay tank principle in preventing scaling.
5. Corrosion in the system in extended operation.
6. Effect of oxidation promoters and inhibitors.
7. Chemical and physical properties of scrubber effluent.

It is recommended also the other research and development work suggested herein be carried out, particularly in regard to testing plant arrangements for intermittent operation and for partial sulfur oxide removal.

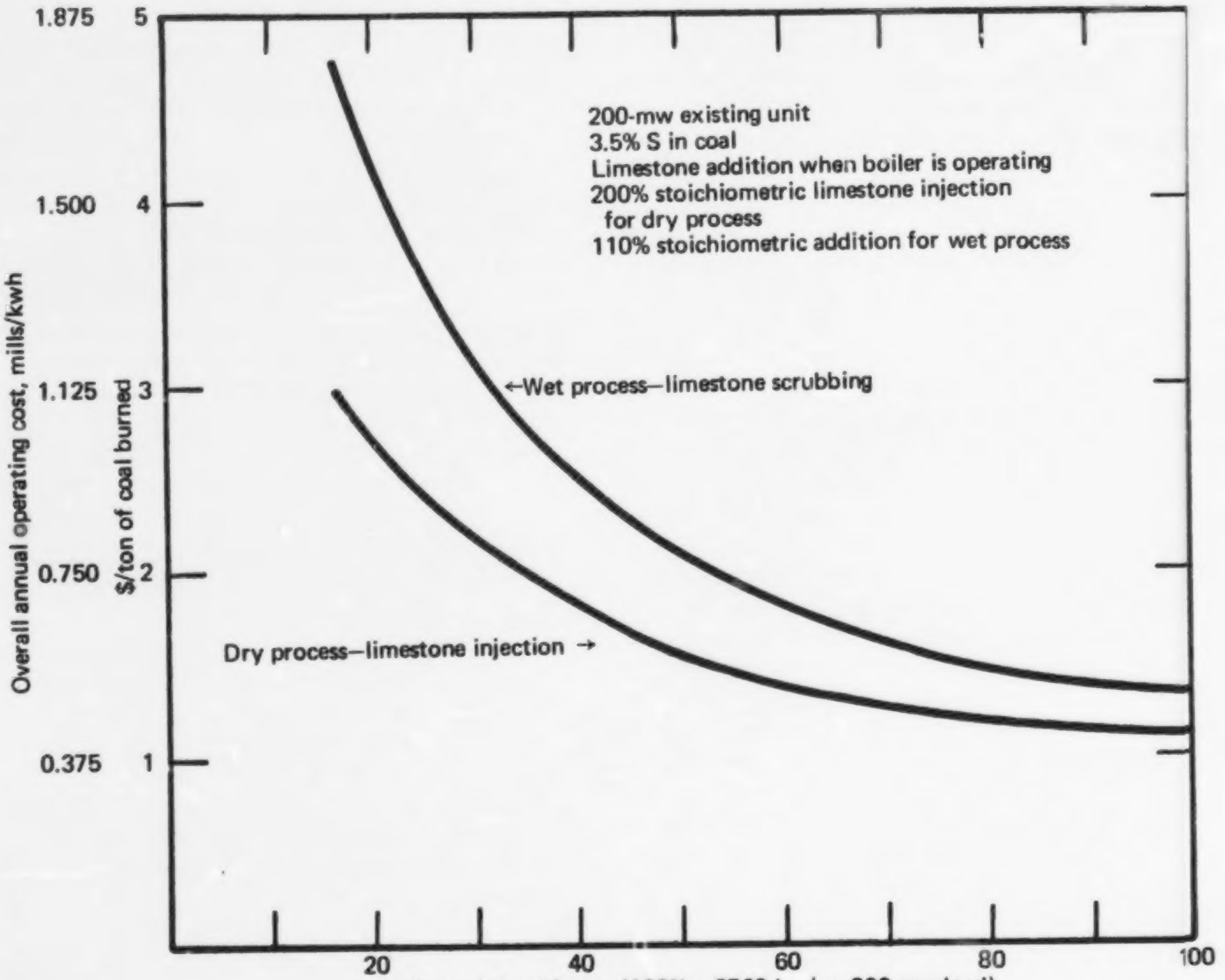

Figure 26. Effect of boiler load factor on wet scrubbing vs dry process operating cost

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Appendix A: Studies of Limestone–Wet Scrubbing

London Power Company (London, England)[1,2]

The London Power Company in the early 1930's developed a process for controlling dust and sulfur oxide emission from the Battersea power station in London. Cyclone collectors were used to remove most of the dust from the boiler exhaust gases. The gases were then passed through vertical steel scrubber elements installed in the main flue where they were contacted in transverse flow with water from the Thames River. Slow corrosion of the steel provided easily and relatively cheaply the iron salts that promoted oxidation of sulfite in the effluent, which was considered necessary to prevent depletion of dissolved oxygen in the river. The effluent was aerated after discharge to further promote oxidation and was finally filtered in gravel beds before discharge back into the river.

The gases passed out of the main flue into the downtakes which discharged into flues leading to the chimneys. Further washing was accomplished on steel scrubber elements installed in the downtakes and in wood grid scrubbers in the flues. At the top of the final scrubbing unit, the gases were contacted with a suspension of finely ground chalk. The chalk (calcium carbonate) was satisfactory as a source of alkalinity (presumably because of reaction with carbon dioxide in the gas to form soluble bicarbonate), but addition of lime was said to cause scaling by reaction with calcium bicarbonate in the river water. Finally, the gases were passed through spray eliminators and discharged to the chimneys.

The effluent from the second and final washing steps was oxidized by aeration but was not filtered because all dust was removed in the first section. The total effluent was mixed with about 15 to 20 times its volume of condenser cooling water before it was returned to the river.

The process removed about 90% of the sulfur dioxide in the flue gas and virtually all of the dust.

The effectiveness of the process as a pollution control method was a controversial matter. A downward path of the washed gas leaving the stack was often observed, especially when the wind speed was low and the gases passed over a relatively cool region. The value of gas washing was questioned since the gases reached the ground much nearer to the chimney than under usual conditions and therefore were less diluted by atmospheric turbulence. From theoretical considerations it was concluded that at least 80% of the sulfur dioxide would have to be removed before maximum ground concentration near the chimney would be reduced. At low wind speeds, observation showed that conditions were worse than predicted.

The large amount of water required and the excessive cooling of the gases make the process unattractive; therefore, no further details will be presented. A detailed account of plant design and operation is given by Hewson, *et al.*

James Howden and Company - Imperial Chemical Industries (London, England)[3]

Also in the early 1930's, the Howden-ICI noneffluent lime scrubbing process was developed in England. The process as envisioned consisted of three main operations: lime treatment, flue gas scrubbing, and separation of solids. The plant requirements were:

1. The lime section–comprising lime or chalk storage, mixing apparatus, and slurry stock.
2. The scrubbing section–consisting of absorption towers and recirculation system.
3. Solids separation section–comprising settlers and filters (or centrifuges), facilities for returning the clarified liquor to the liquor circulation system, and storage for wet solids.

The major problems encountered during process design were (1) finding a nonchoking scrubber packing with high absorption characteristics and suitable scaling resistance, and (2) reducing the scaling potential of calcium sulfate and sulfite in the recirculating solution. The last of these was the most difficult. Theoretical considerations presented by the Howden-ICI researchers on the scaling problem follow.

Addition of lime or chalk to the circulating liquor, which contains dissolved carbon dioxide, was said to result in formation of soluble calcium bicarbonate, the active absorbing agent for sulfur oxides in the wet system. Formation of the bicarbonate from suspended calcium carbonate is continuous because of depletion of dissolved bicarbonate by reaction with sulfur oxides. The calcium sulfite and sulfate formed are relatively insoluble compounds and are precipitated from the circulating liquor. The normal tendency is for these compounds to deposit on absorbing surfaces in the scrubber system with resulting scaling and plugging.

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[1] Rees, R. L. "The Removal of Oxides of Sulfur from Flue Gases." *J. Inst. Fuel, XXV* (148), pp. 350-356 (March 1953).

[2] Hewson, G. W., Pearce, S. L., Pollitt, A., and Rees, R. L. "The Application to the Battersea Power Station of Researches into the Elimination of Noxious Constituents from Flue Gases, and the Treatment of Resulting Effluents." *Soc. Chem. Ind. (London), Chem. Eng. Group, Proc. 15*, pp.67-99 (1933).

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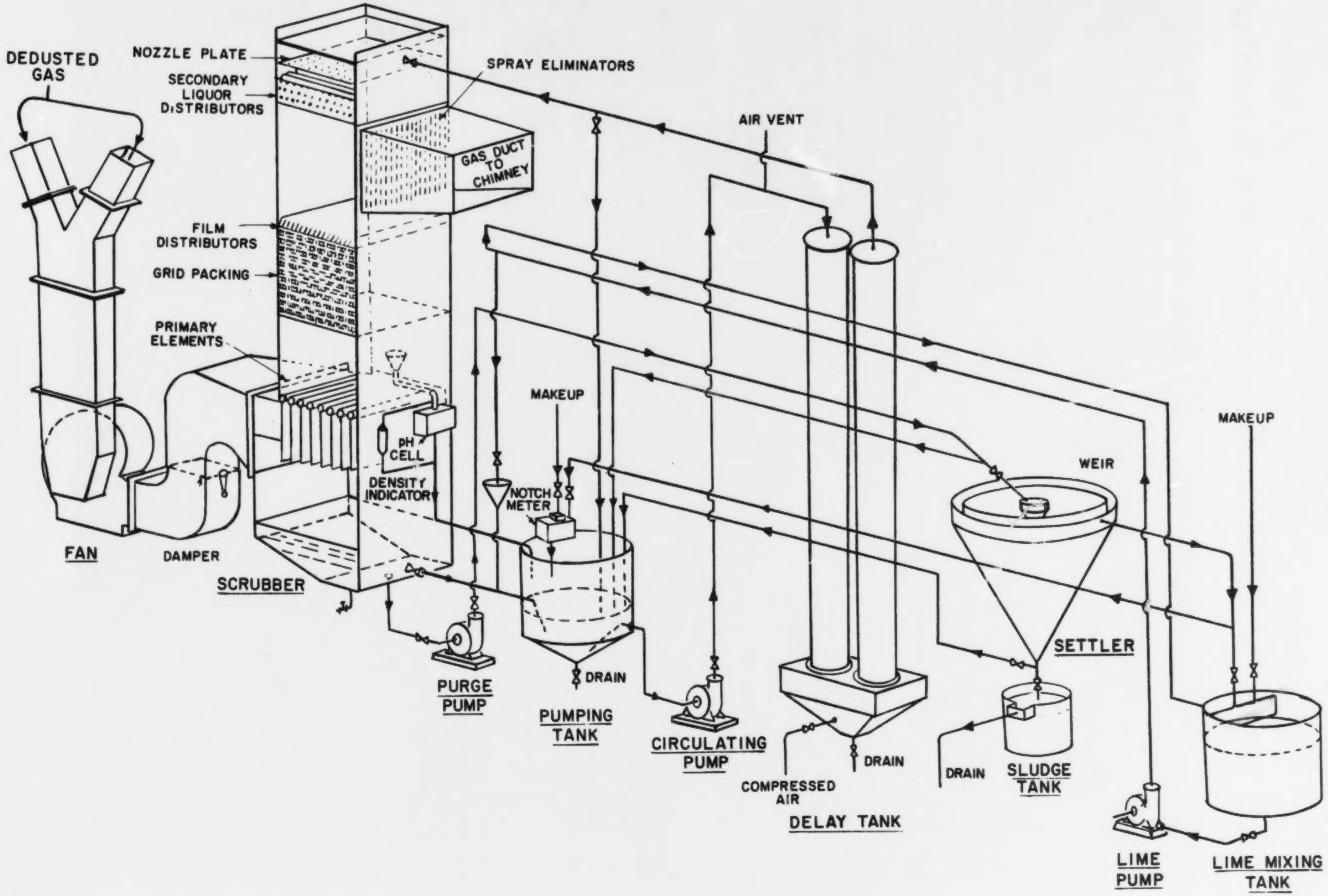
DEDUSTED GAS
NOZZLE PLATE
SECONDARY LIQUOR DISTRIBUTORS
SPRAY ELIMINATORS
AIR VENT
GAS DUCT TO CHIMNEY
FILM DISTRIBUTORS
GRID PACKING
PRIMARY ELEMENTS
MAKEUP
MAKEUP
WEIR
pH CELL
DENSITY INDICATOR
NOTCH METER
FAN
DAMPER
SCRUBBER
PURGE PUMP
DRAIN
PUMPING TANK
CIRCULATING PUMP
COMPRESSED AIR
DRAIN
DELAY TANK
SETTLER
DRAIN
SLUDGE TANK
LIME PUMP
LIME MIXING TANK

Figure A–1. Howden - ICI pilot plant

Calcium sulfite and calcium sulfate form relatively stable supersaturated solutions so that precipitation takes place very slowly, particularly when the solution is only slightly supersaturated. Because of this, it is difficult to prevent scaling by forcing the precipitation to take place away from the scrubber surfaces. Lessing[4] found, however, that this could be accomplished by (1) carrying calcium sulfate and sulfite crystals in the recirculating liquor (minimum content of 3% each) to provide a "harmless" surface for deposition, (2) holding the slurry in a delay tank for a period to allow dissipation of supersaturation on the recirculating crystals, and (3) using "nonscaling" packing (suspended wood plates) in the lower part of the scrubber.

A pilot plant was constructed at Billingham in early 1933 to study the scrubbing process. A schematic diagram of the system is shown in figure A–1. A 4-foot-square, grid-packed absorption tower was used to scrub a portion of the flue gas from a pulverized fuel boiler. Operation with and without dust removal prior to the scrubber was studied. During extended tests, 97 to 99% sulfur dioxide removal and 97 to 98% dust removal were obtained; the absorption rate was found to be controlled almost entirely by the gas film resistance. Design and operating data are shown in table A–1.

Because of the high solids content (up to 20%) in the circulating liquor it was found advisable to: (1) limit liquid flow to 7 feet per second in mild steel pipes, (2) rubber line pipe bends, and (3) use axial flow pumps with rubber-lined casings and rubber-coated impellers.

Provided the pH of the liquor was maintained above a pH of 6.2, corrosion of the scrubber shell was negligible. Although the unit was operated almost continuously for 20 months, the ¼-inch mild steel plate showed no measurable metal loss. Mild steel packing near the gas inlet was appreciably corroded; wood, alloy steel, and brass packings were satisfactory.

Use of the delay tank (2.5 min. retention) plus carrying crystals in suspension virtually eliminated scaling. After 3,500 hours of operation, the packing remained as clean as when it was new.

Lime was used in the first tests but calcium carbonate was used for over 1,000 hours in the later tests. About 30% excess calcium oxide was required when calcium carbonate (finely ground chalk) was used; with lime, the stoichiometric amount was adequate. Precipitated calcium carbonate was also a satisfactory absorbent. The lime or chalk was added continuously to a line just ahead of the delay tank. The increase in pH from 6.2 to 6.8 reduced the solubility of calcium sulfite and sulfate and promoted rapid crystallization. Tests showed that, at 122° F., the solubility of calcium sulfite is 0.637 grain equivalents per gallon at pH 6.2 and 0.518 at 6.8.

Table A–1. Design and Operating Data for the Howden-ICI Pilot Plant

Packing	
No. of primary elements, 3.7 x 2.88 ft	13
No. of wooden grids, 1 in. deep, ¾ in. pitch, 4 ft^2	41
Depth of grid packing, ft	3.5
Total surface area of grid packing, ft^2	1,820
Total surface area of grid packing, plus primary elements, ft^2	2,098
Gas velocities	
At inlet into scrubber (220° F.), ft/sec	13.8
Immediately under packing (220° F.), ft/sec	4.5
At exit through spray eliminators (120° F.), ft/sec	5.9
Pressure drop over whole tower, in. water ga.	1.2
Pressure drop over packing only, in. water ga.	0.8
Analysis of coal fired, %	
Ash	13
S	1.7-2.0
Operating conditions (pulverized fuel firing)	
Equivalent rate of coal fired at 12½% CO_2 lb/hr	1,100
Effective sulfur in coal fired, %	1.7-2.0
Maximum gas rate, c.f.m. at N.T.P.	3,150
Circulating water rate, g.p.m.	260-440
Delay time, min.	2.5-3.5
CO_2 in flue gas, %	12-14
Inlet gas temperature, °F.	220
Exit gas temperature and circulating liquor temperature, °F.	118
Typical concentration of suspended solids, %	
$CaSO_4 \cdot 2H_2O$	3.2
$CaSO_3 \cdot 2H_2O$	2.9
$CaCO_3$	0.4
Ash	7.5
Total	14.0
Alkalinity of liquor to methyl orange at scrubber exit (pH 4.5), gr. equiv./gal	0.2
pH at scrubber exit hopper	6.1
pH at scrubber inlet (head tank)	
Lime addition	6.9
Chalk addition	6.5
Gas analysis, gr./ft^3 at N.T.P.	
Inlet	
Dust	6-13
Total sulfur	0.5-0.8
Outlet	
Dust	0.1-0.2
Total sulfur	0.01-0.02
Alkali consumption, maximum % of theoretical equivalent of SO_2 absorbed	90

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[4] Lessing, R. "The Development of a Process of Flue Gas Washing Without Effluent." *J. Soc. Chem. Ind.*, Trans. and Com., pp. 373-388 (November 1938).

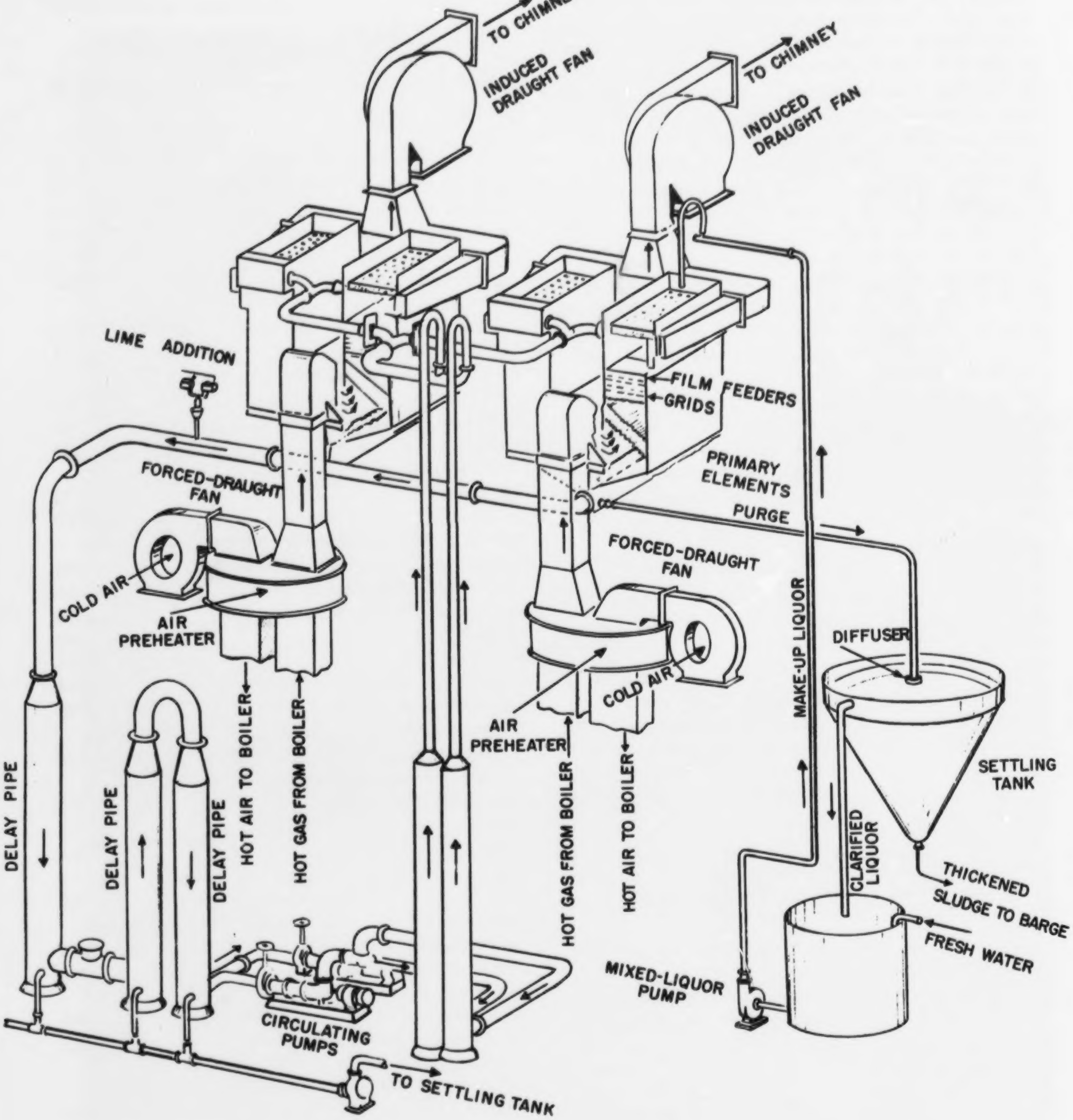
TO CHIMNEY
INDUCED DRAUGHT FAN
TO CHIMNEY
INDUCED DRAUGHT FAN
LIME ADDITION
FILM FEEDERS
GRIDS
PRIMARY ELEMENTS
PURGE
FORCED-DRAUGHT FAN
FORCED-DRAUGHT FAN
COLD AIR
AIR PREHEATER
AIR PREHEATER
COLD AIR
DIFFUSER
MAKE-UP LIQUOR
SETTLING TANK
DELAY PIPE
DELAY PIPE
DELAY PIPE
HOT AIR TO BOILER
HOT GAS FROM BOILER
HOT GAS FROM BOILER
HOT AIR TO BOILER
CLARIFIED LIQUOR
THICKENED SLUDGE TO BARGE
FRESH WATER
MIXED-LIQUOR PUMP
CIRCULATING PUMPS
TO SETTLING TANK

Figure A–2. Howden - ICI unit at Fulham power station

On shutting down, it was necessary to wash the packing to prevent crystal formation on the packing. If through maloperation scaling occurred, chemical cleaning could be accomplished by substituting sodium carbonate for calcium carbonate. The sodium salt reacted with calcium sulfate and sulfite to form calcium carbonate which was dissolved by carbon dioxide in the circulating liquor.

The application of the scrubbing process to a commercial plant, the Fulham power station, was described by Parker and Clarke.[5] A schematic diagram of the istallation is shown in figure A–2. Design conditions are given in table A–2.

Efficiency of the plant was quite good. The sulfur content of the exit gas was as low as 0.006 grain of sulfur per cubic foot (99+% removal) compared with the permissible value of 0.03. Only mere traces of dust (less than 5 microns) were discharged to the atmosphere. Use of the delay tank technique was completely successful in controlling scale formation.

Several maintenance problems developed. Rubber lining was difficult to maintain, particularly with lined pumps. When unlined mild steel pumps were used corrosion and erosion were serious. Use of pumps with shrouded, chrome steel impellers reduced the problem. The original wood packing was reinforced by brass strips and it was found that these brass strips were attacked by dissolved oxygen in the liquor and that dissolved zinc and copper acted as a catalyst to oxidation in the system. This produced corrosion in various parts of the system, and particularly on the mild steel washer tanks. The dissolved zinc and copper also caused pump corrosion through galvanic action. Elimination of the brass strips reduced corrosion to about one-third the original value. Addition of an organic inhibitor (molasses) further reduced the problem.

Tennessee Valley Authority (Muscle Shoals, Alabama)[6]

During early work (1953) by TVA on recovery of sulfur dioxide from power plant stack gases, a pilot plant was constructed, primarily to study an ammonia-scrubbing process. The plant was designed to scrub about 350 standard cubic feet per minute of gas produced by burning approximately 2 pounds per minute of pulverized coal containing 3.5 to 4.0% sulfur; sulfur dioxide concentration in the gas was about 0.3%. The scrubber was constructed of mild steel and was 2 feet in diameter and packed with 8 feet of 2-inch ceramic Raschig rings. Gas-liquid flow was countercurrent with recirculated liquid introduced through a distributor just above the packing. A water spray chamber was provided ahead of the scrubber for cooling of the gas by humidification from 300° to 125° F.

Late in the pilot plant program a test was made for the purpose of investigating the use of a slurry of limestone for removing the sulfur dioxide from combustion gases. The first material tested was a precipitated calcium carbonate (whiting). With slurries containing 10 to 20% solids, less than 25% of the sulfur dioxide was removed even with 50% excess calcium carbonate. The pH of the slurry decreased rapidly from an initial value of 7.3 to about 4 despite the presence of excess absorbent.

With a natural limestone (45.5% CaO) from the middle Tennessee area, a 10% slurry of 95% minus 100-mesh particles removed 89% of the sulfur dioxide. Makeup limestone was added in the stoichiometric quantity

Table A–2. Operating and Design Data from Fulham Sulfur Dioxide Removal Unit

Coal fired (calorific value approx. 10,900 Btu/lb)	31,050 lb/hr
Gas circulated	118,900 cfm
CO_2 content of gas	12.5%
Sulfur content of coal	1.7%
Weight of liquor in circulation	200 tons, approx.
Weight of ash and unburnt carbon caught in the plant	781 lb/hr
Solids formed, assuming 50% oxidation	3,685 lb/hr
Total solids in suspension, with 5% suspended gypsum in recirculating liquor	13%
Rate of circulation of washing liquor	11,500 gpm
Temperature of flue gases entering the plant	240° F.
Temperature of flue gases leaving the plant	124° F.
Humidity of gases leaving the washer	100%
Makeup water required	32.7 gpm
Rate of flow of liquor to the settling plant	496 gpm
Rate of flow of returned or clarified liquor from the settling plant	445 gpm
Complete weight of one unit in working order, exclusive of delay pipes	300 tons
Total cross section of a washer	400 ft^2
Number of cells/boiler	16
Maximum permissible gas flow through one scrubber at 240° F.	118,900 cfm
Total capacity of the delay tanks	6,300 ft^3
Delay period	3.5 min
Final volume of gas/boiler leaving the washer at 124° F.	105,000 cfm
Lime requirements, assuming an excess of 10% and a purity of lime of 92% Cao	1,100 lb/hr

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[5] Parker, W. C., and Clarke, H. *J. Inst. Civ. Engrs. 9*, pp. 28-36 (1937-1938).

[6] Tennessee Valley Authority. Progress Report Assembly No. 47A, February 1954 to September 1955, pp. 72-74 (unpublished).

required to react with all of the sulfur dioxide in the gas. Results from tests with this material are shown below:

Recirculation rate			
Gal/min	Gal/(min)(ft^2)	pH	SO$_2$ removal, %
5	1.6	6.5	81.5
10	3.2	6.6	89.0

The above tests were made with gas that had been cooled by humidification to about 125° F. In tests without cooling, the gas entered the scrubber at about 300° F. and removal of sulfur dioxide decreased to 54%; pH of the slurry was 5.0. Use of large excesses of limestone did not result in improved absorption nor increased pH. The slurry temperature was only 5° higher (130° vs 125°) than when humidified gas was used and, therefore, probably was not responible for the poor removal of sulfur dioxide. When the humidifier was used, most of the sulfur trioxide was removed in the humidifier effluent water.

In additional tests with redwood grids rather than rings as packing in the scrubber, two additional limestones and one dolomite were tested; the supply of stone used in the earlier tests had been nearly exhausted. None of these materials were satisfactory even when used in excess. Sulfur dioxide removal efficiency varied from 6 to 15% and pH of the slurry in each case was about 4.5.

A series of bench-scale exploratory tests was made in an effort to determine why the limestones used in the last pilot plant tests were ineffective in removing sulfur dioxide. Measurements were made of the amount of sulfur dioxide removed from simulated flue gas when the gas was bubbled through 10% slurries of various stones. The different limestones varied in reactivity but the stone which gave good results in the early pilot plant tests was not appreciably more reactive than the other types tested. It was concluded that absorption was more closely related to the ratio of gas rate and liquid volumes than to the type of limestone used.

The low-priority, exploratory program on limestone scrubbing was terminated so that work could continue on the more promising ammonia process.

Scientific Research Institute (Soviet Union)[7]

In studies by Chertkov and coworkers at the Scientific Research Institute in Russia, mass transfer coefficients during absorption of sulfur dioxide from gases by lime suspensions were calculated from data obtained during tests in a pilot-scale (2,500 c.f.m.) grid-packed absorber. Absorption from gases containing 0.2 to 0.3% sulfur dioxide and about 6.5 grains per cubic foot of fly ash varied from 96 to 99% when the pH of the circulating liquor was held at 6.1 to 6.2. The mass transfer coefficient was directly proportional to the gas velocity for these conditions, under which the equilibrium vapor pressure of sulfur dioxide was nearly zero at the scrubber outlet.

With decrease in pH the equilibrium pressure increased sharply from zero at pH 8.0 (pure calcium sulfite) to about 2.3 millimeters Hg, the same as the inlet gas, at a pH of about 4.5. The mass transfer coefficient decreased from 150 to 120 pounds per square foot per hour per atmosphere as the pH of the absorbing solution decreased from 6.3 to 5.3; gas velocity was held constant. At pH values higher than 6, the liquid phase resistance was practically zero and the general coefficient of mass transfer could be equated to the partial coefficient of the gas phase.

The effect of solids content of the scrubbing solution on the transfer coefficient at various gas flow rates is shown in figure A–3. The strong influence of solids content was in part a function of slurry viscosity but also was related to the type of packing. Use of finely divided packing instead of the grid type used would have reduced the effect of solids on the transfer coefficient.

Wisconsin Electric Power Company and Universal Oil Products, Inc., Air Correction Division[8]

A study of reducing sulfur dioxide emissions from power plant stack gas by lime scrubbing was initiated by Wisconsin Electric Power Company in 1964; the Air Correction Division of Universal Oil Products Company, the manufacturer of the Turbulent Contact Absorber (TCA) used in the project, contributed to the program.

A pilot TCA scrubber was installed on a 120-megawatt boiler (Combustion Engineering design) at Wisconsin Electric's Oak Creek station. The scrubber, constructed of fiber glass-reinforced polyester, was 14 inches in diameter and 25 feet in overall height with provision for four scrubbing sections each containing a bed of 1½-inch-diameter hollow polypropylene spheres. Gas entered the bottom of the scrubber and slurry was sprayed into the top. A spray nozzle was installed near the gas inlet to adiabatically cool the incoming gas and to assist in removal of fly ash. A chevron-type mist eliminator was provided at the top of the scrubber. A diagram of the pilot plant equipment is shown in figure A–4.

The absorbent slurry was prepared in a feed tank and metered through a rotameter to the slurry recirculation pump. Dust loadings, sulfur dioxide samples, and pitot tube traverses were taken at inlet and outlet sample points.

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[7] Chertkov, B. A. "Coefficients of Mass Transfer During Absorption of SO$_2$ from Gases by Lime Suspension." *Khim. Prom.* No. 7, pp. 533-36 (1962).

[8] Pollock, W. A., Tomany, J. P., and Frieling, G. "Flue-Gas Scrubber." *Mech. Eng. 89* (8), pp. 21-25 (August 1967).

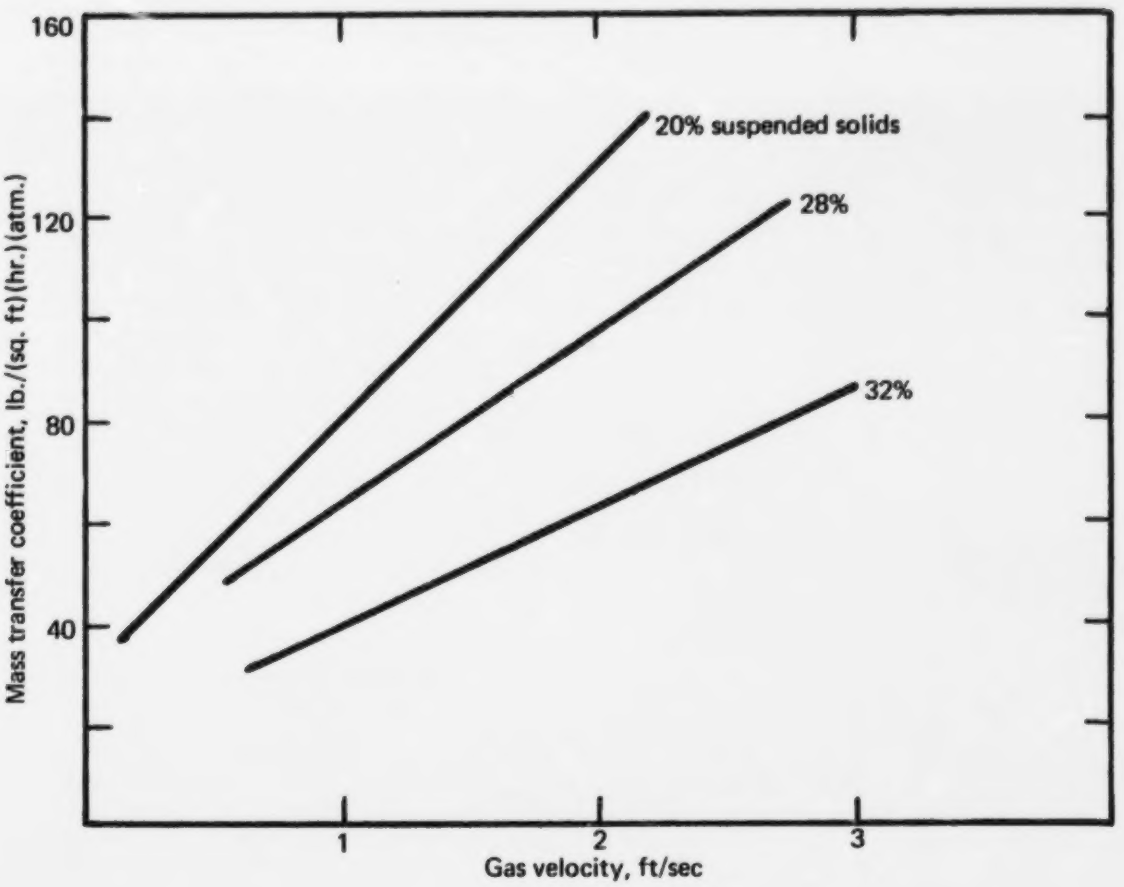

Figure A–3. Effect of suspended solids in lime slurry on mass transfer coefficient for SO_2 absorption

The majority of tests for sulfur dioxide absorption were made with two stages of scrubbing. Both sodium and calcium carbonate (limestone) were used as absorbents.

The absorbent efficiency with sodium carbonate ranged from 89.6 to 94.4% as the liquid rate was increased from 20 to 40 gallons per minute per square foot, at a constant gas rate of 750 feet per minute. Efficiencies with limestone were 79.8 and 83.6% at 30 and 50 gallons per minute per square foot, respectively, at a gas rate of 750 feet per minute. With one stage of scrubbing, about 20% of the nitrogen oxides was removed; efficiency was independent of the absorbent used.

Inlet dust loadings ranged from 1.5 to 2.5 grains per standard cubic foot. At a gas velocity of 920 feet per minute with two-stage operation, 96.9, 98.9, and 99.5% of the dust was removed at liquid rates of 20, 25, and 30 gallons per minute per square foot, respectively. With one scrubber stage, dust collection efficiency was about 5% lower than with two stages under comparable conditions. Pressure drop was approximately 3 inches of water with one stage and 6 inches with two stages, at a liquid rate of 40 and a gas rate of 750.

During operation of the pilot scrubber, inspection of the interior surfaces indicated only a very thin film of calcium carbonate and sulfate which could be easily wiped off. No problems with plugging were encountered. The spheres showed no signs of attrition or excessive wear after more than 500 hours of operation.

The process visualized in these tests included injection of limestone into the boiler to calcine the absorbent and increase its reactivity. The injection step was tested but not at the Oak Creek station where the scrubber tests were made.

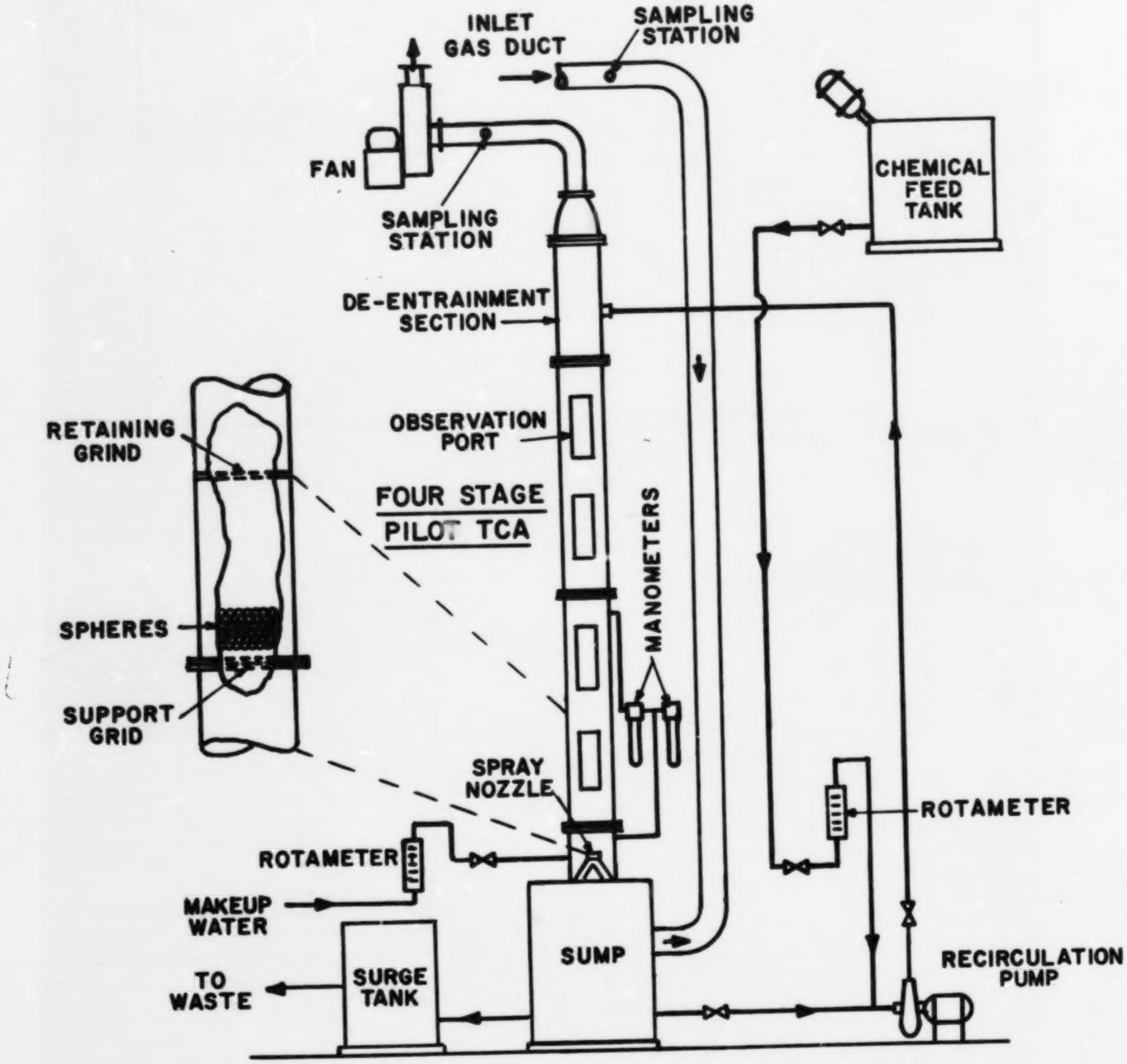
INLET GAS DUCT
SAMPLING STATION
FAN
SAMPLING STATION
CHEMICAL FEED TANK
DE-ENTRAINMENT SECTION
RETAINING GRIND
OBSERVATION PORT
FOUR STAGE PILOT TCA
MANOMETERS
SPHERES
SUPPORT GRID
SPRAY NOZZLE
ROTAMETER
ROTAMETER
MAKEUP WATER
SUMP
SURGE TANK
TO WASTE
RECIRCULATION PUMP

Figure A–4. Scrubber arrangement in Wisconsin power tests

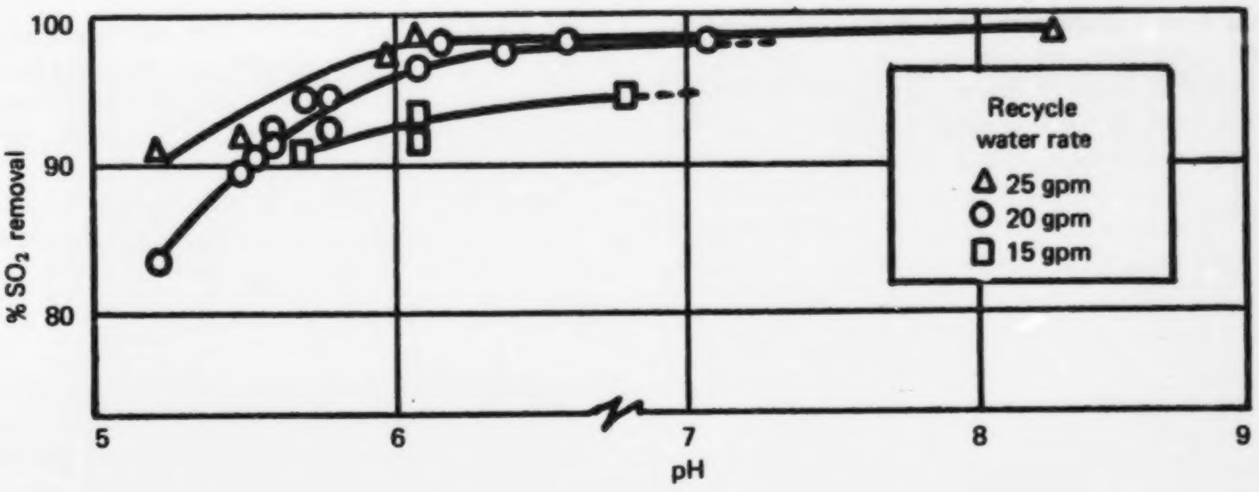

Figure A–5. Percent SO_2 removal vs pH of scrubber effluent

Combustion Engineering, Inc., and Detroit Edison Company,[9, 10]

A system for removal of fly ash and sulfur dioxide from power plant stack gas by lime scrubbing has been studied by Combustion Engineering and Detroit Edison; the work was reported in early 1967. The process is similar to that tested by Wisconsin Electric Power; limestone or dolomite is injected into the boiler to improve absorbent efficiency by calcination, followed by wet scrubbing to remove ash and lime from the gas and to promote reaction of sulfur oxides with the lime.

A small pilot plant was first constructed and tested. Controlled amounts of sulfur dioxide, calcined or uncalcined dolomite, and fly ash were added to flue gas (at 250° F.) from a natural gas-fired boiler and the mixture (1,000 c.f.m.) was passed through a scrubber. With a dust loading of 6.6 grains per standard cubic foot, a spray water rate of 20 gallons per minute resulted in 98% dust-collection efficiency and 99% sulfur dioxide removal at a pH of 9.2 (125% stoichiometric calcined dolomite). Only about half this efficiency was obtained with raw dolomite.

The next phase of the study was conducted jointly by Combustion Engineering and Detroit Edison at the latter's St. Clair power plant. The objective was to determine if the pilot unit data could be confirmed on a commercial-size unit. Dolomite was introduced on a full-scale basis to one furnace of a 325-megawatt twin furnace unit (tangentially fired). The dolomite was ground in one of the coal pulverizers and injected through the top burners of the boiler.

Approximately 3,000 actual cubic feet per minute of exhaust gas (about 1% of total) was withdrawn after the air heater and passed through a fluid bed (glass marble) scrubber (National Dust Collector Hydro-Filter). The gas flowed upward together with a spray of recirculated slurry through the marble bed. The clean gas passed through a demister to the stack and the slurry drained back through the bed and to a settling tank.

Continuous tests were made for periods ranging up to 5 days. No scaling or plugging problems were reported. Sulfur dioxide removal efficiency dropped 2 to 5% when the angle of injection into the boiler was changed so as to expose the dolomite to higher temperature. Efficiency increased from about 90% to 98% as the liquid to gas ratio was increased from 7 to 12 gallons per 1,000 cubic feet. The effect of scrubber effluent pH is shown in figure A-5; a pH of at least 6 was required for good absorption. It was found that the pH increased when the effluent was held in a mixing tank for a period prior to discharge to the settling tank. The sludge from the settling tank contained about equal amounts of calcium sulfate and calcium sulfite and the liquid phase was nearly saturated with magnesium sulfate. About 20% of the nitrogen oxides in the flue gas was removed in the scrubber.

From a tabulation of data from various runs, optimum

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[9] Plumley, A. L., Whiddon, O. D., Shutko, F. W., and Jonakin, J. "Removal of SO_2 and Dust from Stack Gases." Paper presented at American Power Conference, Chicago, Illinois, April 25-27, 1967.

[10] Plumley, A. L., Jonakin, J., Martin, J. R., and Singer, J. G. "Removal of SO_2 and Dust from Stack Gases." *Combustion 40* (1), pp. 16-23 (July 1968).

conditions were: scrubber pH, 6.5; slurry rate, 7.6 gallons per 1,000 cubic feet; gas velocity, 530 feet per minute; pressure drop, 6 inches of water; dolomite feed, 110% of stoichiometric. Under these conditions, the system removed 98% of the sulfur dioxide and 99.5% of the dust. A schematic diagram of the injection-scrubbing system is shown in figure A-6. (Although a gas reheating system is shown, this equipment was not included in the test facility.)

In a later paper by Plumley et al. (Combustion Engineering), commercial installations of the scrubbing system are described. The first unit was installed at Unit No. 2 in the Union Electric Company's Meramec station (St. Louis) and was started up in November 1968. The 140-megawatt boiler on which the unit was installed is tangentially fired and burns pulverized midwestern bituminous coal containing about 3.4% sulfur and 9% ash. The scrubbers were installed above the precipitators, which were blanked off and bypassed. Limestone is fed to an existing mill at a rate equivalent to 110% of stoichiometric. A vertical baffle was installed in the bunker to provide storage space for limestone. Induced-draft fans were modified for increased draft loss by installing new wheels to operate at 880 r.p.m. instead of 700 r.p.m.; larger motors were required.

The process is also being installed at the Lawrence, Kansas, plant of the Kansas Power and Light Company, on Units 4 and 5. Unit 4, a 125-megawatt tangentially fired boiler, has been operating since 1960; the scrubber system was completed late in 1968. Unit 5 is a 420-megawatt installation now under construction, with operation planned for early 1971.

Tests were conducted at the Lawrence Unit 4 to determine the feasibility of combining limestone with coal prior to grinding. Results showed that:

1. Coal and limestone segregated less than 5% in the bunker.
2. Pulverizer power requirement for the mixture was about the same as for coal.
3. Furnace operation was improved by firing the mix.
4. More than 30% of the sulfur was removed by in-furnace reactions and by pyrite rejection.

The limestone and coal will be premised in Unit 4 system but Unit 5 will be fitted with separate injection facilities. A system flow diagram for Unit 5 is shown in Figure A-7.

The Meramec and Lawrence installations are guaranteed to remove 99% of the dust and to reduce sulfur dioxide to a level equivalent to burning coal containing 0.5% sulfur. The Meramec unit has separate additive feed, uses a liquid-couple stack gas reheating circuit, and has a clarifier for concentration of solids. The two Kansas Power and Light units will have hot water reheaters and the fly ash pond will be used for slurry clarification.

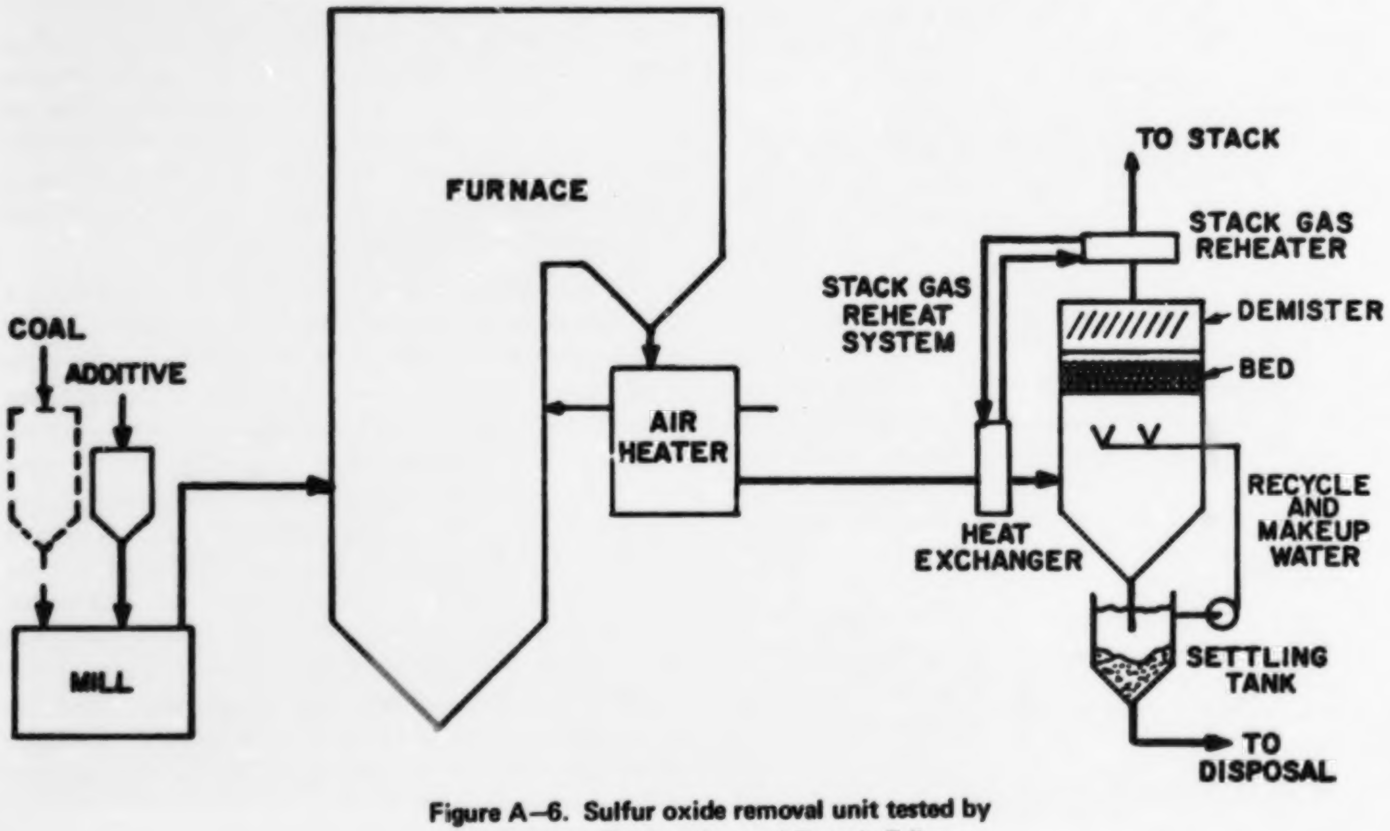

Figure A–6. Sulfur oxide removal unit tested by Combustion Engineering and Detroit Edison

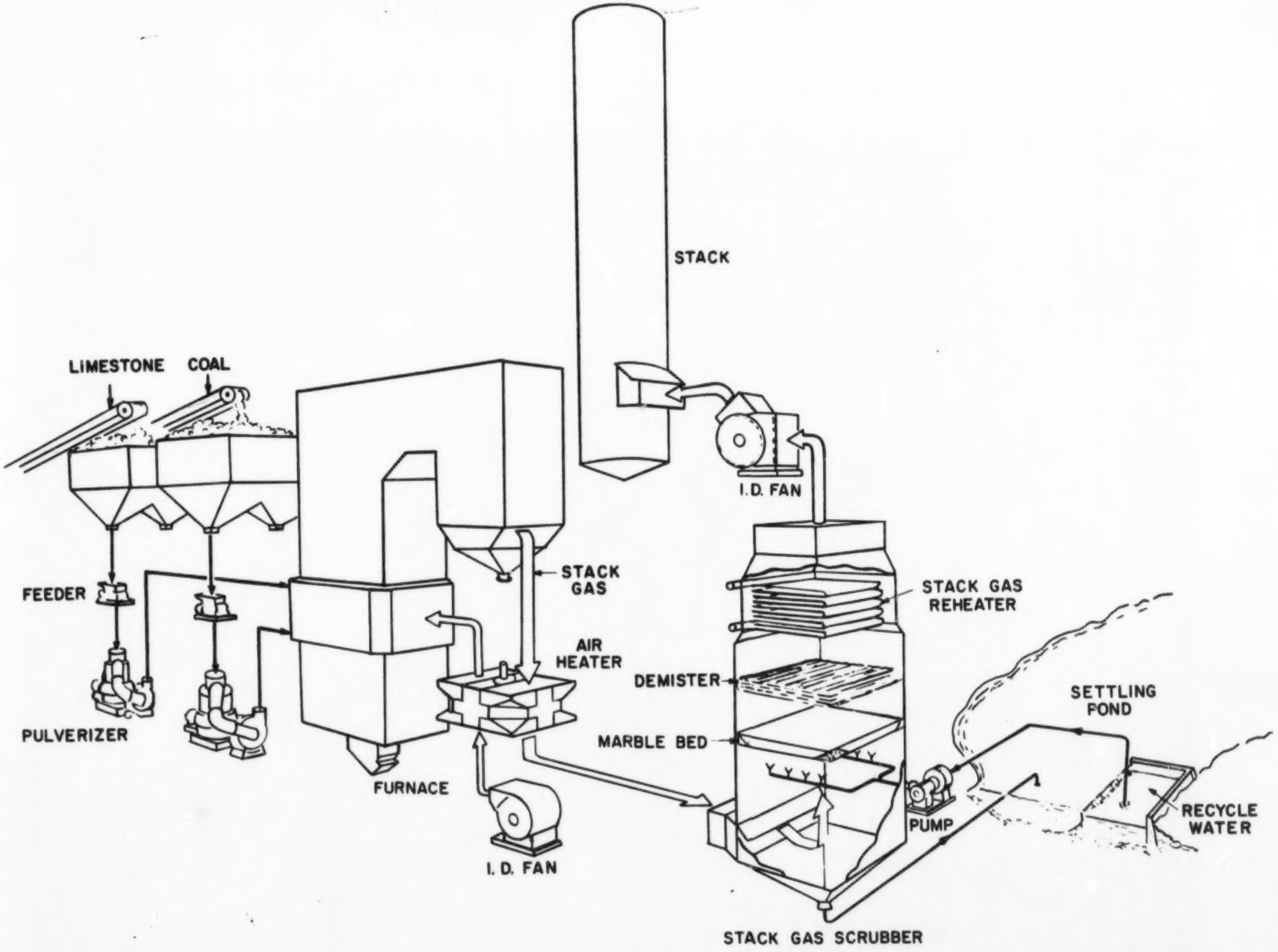
STACK
LIMESTONE
COAL
I.D. FAN
FEEDER
STACK GAS
STACK GAS REHEATER
AIR HEATER
DEMISTER
SETTLING POND
PULVERIZER
MARBLE BED
FURNACE
PUMP
RECYCLE WATER
I. D. FAN
STACK GAS SCRUBBER

Figure A–7. Lime scrubbing system to be installed on boiler unit 5 of Kansas Power and Light Company

Appendix B: Water Quality Considerations

The feasibility of releasing dissolved or suspended calcium salts into watercourses depends on how much the content of these impurities in the water course is increased, which in turn depends on the amount of the particular pollutant released in the effluent, the content in the watercourse before influx of the effluent, and the water volume flow of the watercourse. Various regulatory agencies and advisory groups have set or recommended limits above which the concentration of an impurity in a watercourse should not be increased. Such limits for the solids and dissolved salts released from a limestone–wet scrubbing process are summarized herein. Also presented is an analysis, based on these limits, of the degree of pollution by limestone–wet scrubbing; for this it is assumed that the power plant involved is a 200-megawatt unit, the coal contains 3.5% sulfur, the scrubber is designed for 95% removal of sulfur oxides, the limestone contains no magnesium, and 110% of the stoichiometric amount of absorbent is used. The following quantities are involved.

Total solids to pond	
$CaSO_4$	48,800 tons/yr
$CaSO_3$	27,400
$CaCO_3$	6,720
Inert	3,860
Fly ash	52,000
Water of hydration	21,000
	160,000 tons/yr
or	40,000 lb/hr
Free water (10% slurry)	360,000 lb/hr
Dissolved solids to watercourse	
$CaSO_4$	890 lb/hr
$CaSO_3$	14
$CaCO_3$	7

1. **Suspended solids**

Most regulatory agencies do not place quantitive limits on either suspended solids or turbidity for raw water used as a source for domestic water, but do generally specify that concentrations shall not be sufficiently high as to be objectionable or interfere ith normal treatment processes. Both turbidity and suspended solids can be reduced to acceptable limits for most purposes by conventional treatment methods; however, high concentrations increase the cost of processing the water by increasing chemical requirements and the volume of sludges to be disposed of. In some cases head losses through the filters are increased with a resulting increase in the frequency of filter backwashing.

Settleable suspended solids often blanket stream bottoms, killing fish eggs and young and destroying much of the normal benthic biota which serve as food organisms required for propagation and growth of fish. Turbidity in streams also interferes with light penetration and limits or diminishes the photosynthetic effects necessary for the primary productivity of fish-food organisms. The Committee on Water Quality Criteria of the Federal Water Pollution Control Administration[1] recommends a limit of 50 Jackson Turbidity units in warm water streams. TVA biological surveys indicate that ash discharge from power plant ash-settling ponds is not sufficient to adversely affect the normal benthic environment of the river channel or otherwise cause water quality problems. Whether or not this situation can be maintained with limestone–wet scrubbing will depend on settling characteristics of the lime-calcium sulfate-ash slurry and the degree to which pond area is increased to handle the increased solids load.

2. **Total dissolved solids (2,600 mg./l. in effluent from 200-mw. unit, from quantities given in above tabulation)**

The following information is quoted from *Water Quality Criteria*, issued by the Resources Agency of California.[2]

(a) Domestic Water Supplies. The 1962 USPHS Drinking Water Standards specify that the total dissolved solids should not exceed 500 mg./l. if more suitable supplies are, or can be made available. This limit was set primarily on the basis of taste thresholds. The 1958 WHO International Standards set the "permissible limit" at 500 mg./l. and the "excessive limit" at 1500 mg./l., but no "maximum allowable limit" is given. The 1961 WHO European Standards do not include limits for total dissolved solids. It is generally agreed that the salt concentration of good, palatable water should not exceed 500 mg./l.; however, higher concentrations may be consumed without harmful physiological effects and may indeed even be more beneficial. Each water with a total salt concentration over 1000 mg./l. should be judged on the basis of the local situation, alternative supplies, and the reaction of the local population.

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[1] Federal Water Pollution Control Administration. *Water Quality Criteria* (Report of the National Technical Advisory Committee to the Secretary of the Interior). U.S. Government Printing Office, Washington, D. C. (April 1, 1968).

[2] McKee, J. E., and Wolf, H. W. *Water Quality Criteria*. The Resources Agency of California, State Water Quality Control Board, Publication No. 3-A (1963).

(e) Fish and Aquatic Life. It has been reported that among inland waters in the United States supporting a good mixed fish fauna, about 5% have a dissolved-solids concentration under 72 mg./l.; about 50% under 169 mg./l., and about 95% under 400 mg./l.

Water quality objectives adopted by TVA for surface water in the Tennessee River basin include a limit of 500 milligrams per liter of dissolved solids, which would require a dilution of about 5:1 for the 2600 milligrams per liter of dissolved solids from the 200-megawatt unit—easily attainable with even a relatively small receiving stream for the effluent flow involved (about 1.5 cu. ft./sec. for the 200-mw. unit). For larger power plants, say a total of 2000 megawatts at one site, the required dilution of 50:1 would also be practicable in most situations. The flow of diluting water required would be 75 cubic feet per second; in comparison, the approximate average flow past various TVA power plants is 51,000 cubic feet per second for the Colbert plant (on the Tennessee River), 18,000 cubic feet per second at Gallatin (Cumberland River), and 4,300 at Bull Run (Clinch River).

3. **pH (no lower than 6.2)**

TVA has proposed upper and lower pH limits for the waters of the Tennessee Valley of 8.5 and 6.5, respectively. The pH values required for industrial water supplies are covered in some detail in the FWPCA Water Quality Criteria report. The proposed TVA criteria should be adequate for all of these but a few extreme cases. For limestone–wet scrubbing, the scrubber effluent should have a pH no lower than 6.2 for good scrubber operation; since the pH of watercourses normally is considerably higher than this (7.5 in the Tennessee River), dilution should be adequate, even by small watercourses, to reach the minimum pH level of 6.5. Any detrimental effects of pH should be confined to a small mixing zone in the vicinity of the waste outfall.

4. **Calcium (Ca^{++}, 756 mg./1.; 1,860 mg./1. expressed as $CaCO_3$ hardness)**

Calcium in water supplies has been suspected of producing undesirable physiological reactions but no definite relationship has yet been established. The major concern is related more to its deleterious effects in uses such as washing, bathing, and laundering and to the problem of incrustations on cooking utensils and water heaters. Some industrial processes require a very soft or low calcium water. Costs of conditioning water for such industries is directly related to source concentration.

Thus the major detrimental effect of calcium hardness is economic. TVA has proposed the following limitation on water hardness:

> There shall be no substance added to the waters that will increase the hardness to such an extent as to appreciably impair usefulness as a source of water supply or interfere with other reasonable and necessary uses of the water.

Rivers in the Tennessee Valley on which steam plants are located generally range from 50 to 125 milligrams per liter hardness as calcium carbonate. In general, a concentration of 100 milligrams per liter would not be objectionable for most uses if this concentration were not significantly greater than the normal range for the area. Thus with a background concentration of 88 milligrams per liter hardness (median between 50 and 150) and an effluent concentration of 1,860 milligrams per liter, a dilution ratio of about 150:1 would be required for a 100-milligram-per-liter limit. This is quite easily obtained for the 200-megawatt plant. However, other sources of hardness (such as fly ash) could elevate the effluent concentration. Moreover, a 2,000-megawatt plant would require a 2,250-cubic-foot-per-second flow of diluting water, or over 50% of the average (Clinch River) flow. This would likely require a pipeline and possibly a diffuser into the river to prevent delay in mixing and consequent high concentration along the near river bank for some distance downstream.

5. **Calcium Sulfate (2,500 mg./l.)**

No limits have been placed on calcium sulfate by regulatory agencies. The following summary of beneficial uses and reported acceptable concentrations is quoted from McKee and Wolf.

(a) Domestic Water Supplies. High calcium sulfate concentrations in water are disadvantageous for most household uses, but for drinking purposes, 300 mg/l or more is not harmful. The taste threshold of calcium sulfate has been reported to be 250 to 900 mg/l. Calcium sulfate significantly increases plumbosolvency; raising the calcium sulfate concentration from 25 to 250 mg/l increased by 10 percent the amount of lead dissolved from lead pipes.

(b) In general, calcium sulfate is beneficial in the brewing industry, for it helps to maintain the acidity of the wort and therefore causes more complete coagulation of albuminous matter. It also reduces the solubility of the bitter substances of the hop.

(c) Irrigation. Gypsum in irrigation waters improves or restores the permeability and tilth of soil having an unfavorable sodium ratio.

(d) Stock and Wildlife Watering. A saturated solution of calcium sulfate, used as the sole source of drinking water for rats, permitted satisfactory growth. A concentration of 2400 mg/l of calcium sulfate permitted normal growth and reproduction among rats.

(e) Fish and Other Aquatic Life. Trama reported that a saturated solution of calcium sulfate at 20° C did not produce significant mortalities among bluegills; but a later report from the same organization indicated that this concentration was the 96-hour TL_m at 18-20° C in soft water for the bluegill sunfish. It was also reported that 3200 mg/l caused a 50 percent reduction in the rate of growth of the diatom, *Navicula seminulum*. Using highly turbid water at 21-25° C and the mosquitofish (*Gambusia affinis*) as the test animal, Wallen *et al.* indicated that the 96-hour TL_m was greater than 56,000 mg/l, a concentration much larger than the solubility.

TVA has proposed no specific limits for calcium sulfate; however, the concentration is limited by the allowable sulfate concentration and the permissible added calcium hardness.

6. Sulfates ($SO_4^{=}$, 1,770 mg./l.)

McKee and Wolf discuss sulfate content as follows.

(a) Domestic Water Supplies. The 1962 Drinking Water Standards of the USPHS recommend that sulfates not exceed 250 mg/l, except where a more suitable supply is not available. This limit does not appear to be based on taste or physiological effects other than a laxative action toward new users. Public water supplies with sulfate contents above this limit are commonly and constantly used without adverse effects. The 1958 WHO International Standards established a "permissible limit" of 200 mg/l and an "excessive limit" of 400 mg/l, but set no maximum allowable limit. The 1961 WHO European Standards include a recommended limit of 250 mg/l for sulfates.

Korotchenok reported on heavily mineralized drinking waters in western Turkmenia (USSR), noting that no outbreaks of disease have been ascribed to waters in which the content of sulfate did not exceed 1295 mg/l. In British Somaliland well waters used for human consumption contain very high sulfates, many having 2000-3000 mg/l. One village is using water of 4400 mg/l sulfates. A survey in North Dakota indicated that water containing less than 600 mg/l of sulfates is usually safe. In reviewing the literature, Moore quotes Sollman to the effect that concentration of 1000 mg/l of sulfates in water is harmless. A cathartic dose is 1.0 to 2.0 grams, or a liter of water containing 1000 to 2000 mg/l of sulfates.

Sulfates appear to have no detrimental effect on the corrosion of brass fittings in domestic water systems, nor do concentrations less than 200 mg/l increase plumbosolvency.

Whipple is quoted by Moore to the effect that the taste thresholds of sulfate salts were as follows:

Sodium sulfate	200-500 mg/l
Calcium sulfate	250-900 mg/l
Magnesium sulfate	400-600 mg/l

(b) Industrial Water Supplies. Limiting or threshold concentrations, or optimum ranges, for sulfates are assembled below:

	Limits of recommended values, mg/l			
Industrial process	**$SO_4^{=}$**	**$CaSO_4$**	**$MgSO_4$**	**Na_2SO_4**
Brewing	–	100-500	100-200	100
Carb beverages	250	–	–	–
Concrete corrosion	25	–	–	–
Ice making	–	300	130-300	300
Milk industry	60	–	–	–
Photo processes	–	100	–	–
Sugar making	20	–	–	–
Textiles	100	–	–	–

TVA has proposed a limit of 150 milligrams per liter for sulfates to provide a reasonably satisfactory water for both domestic and industrial use. Normal water treatment has little effect on sulfate concentration. Assuming a normal background of 15 milligrams per liter sulfate and an effluent of 1,800 milligrams per liter, dilution requirement would be approximately 12 to 1. Again this should be no problem on major streams.

7. Sulfite ($SO_3^{=}$, 27 mg./l.)

The major concern in regard to high sulfite concentration is oxidation to sulfate, thus lowering the dissolved oxygen content of the receiving water. If we assume 95% saturation of the effluent with dissolved oxygen, a temperature of 85° F (29.4° C), and complete conversion of sulfite to sulfate prior to discharge to the receiving stream, the dissolved oxygen content of the effluent would be reduced to 1.6 milligrams per liter. Assuming 95% of saturation in a 77° F receiving stream (7 mg./l. of dissolved oxygen), no significant reduction in the dissolved oxygen of the stream would occur and a dilution ratio of only 1.7:1 would be required to maintain a

> minimum dissolved oxygen content of 5.0 milligrams per liter. No problem should arise from such discharges.

As noted earlier, this treatment of the dissolved solids problem is based on the assumed use of calcitic limestone. There may be also a considerable dissolved content of magnesium sulfite and magnesium sulfate if dolomite is used as the absorbent. Although data indicate little or no reaction of magnesium oxide in the boiler, reaction can occur in the scrubber and also in the disposal pond. Tests at TVA showed that solid magnesium oxide in the pond could react with calcium sulfate and atmospheric carbon dioxide to form magnesium sulfate. In tests with a mixture of 20 grams MgO, 80 grams $CaSO_4 \cdot 2H_2O$, and 300 milliliters H_2O, contact with carbon dioxide resulted in 70% dissolution of the magnesium and 65% of the sulfate.

The effect of dissolved magnesium has been treated by McKee and Wolf.

> (a) Domestic Water Supplies. Magnesium is an essential mineral element for human beings; the daily requirement of magnesium is about 0.7 gram. Magnesium is considered relatively non-toxic to man and not a public health hazard because, before toxic concentrations are reached in water, the taste becomes quite unpleasant. At high concentrations, magnesium salts have a laxative effect, particularly upon new users, although the human body can develop a tolerance to magnesium over a period of time.
>
> The 1946 USPHS Drinking Water Standards recommended a limit of 125 mg/l but there is no limit in the 1962 standards. The 1958 WHO International Standards have a "permissible limit" of 50 mg/l and an "excessive limit" of 150 mg/l, but no maximum allowable concentration. The 1961 WHO European Standards have a recommended limit of 125 mg/l, but if the sulfate exceeds 250 mg/l, the magnesium is limited to 30 mg/l.
>
> The taste threshold for magnesium (in $MgSO_4$) has been reported as 100 mg/l and for the average individual it is given as about 500 mg/l.
>
> The negative correlation between hardness in water and cardiovascular disease does not appear to hold for magnesium as it does for calcium; yet one investigator reports the favorable use of magnesium sulfate to treat such cases and claims that magnesium, rather than calcium, is the beneficial element in reducing cardiovascular attacks.
>
> (e) Fish and Other Aquatic Life. The relative concentrations of magnesium and calcium in water may be one factor controlling the distribution of certain crustacean fishfood organisms, such as copepods, in streams. Hart et al. cite a report that among U.S. waters supporting a good fish fauna, ordinarily 5 percent have less than 3.5 mg/l of magnesium; 50 percent have less than 7 mg/l; and 95 percent have less than 14 mg/l.
>
> Magnesium chloride and nitrate can be toxic to fish in distilled water or tap water at concentrations between 100 and 400 mg/l as magnesium. However, magnesium chloride, nitrate, and sulfate, at concentrations between 1000 and 3000 mg/l as magnesium have been tolerated for 2-11 days. Some fresh-water fish have been found in very saline lake water containing over 1000 mg/l of magnesium as well as additional sodium and calcium salts.

TVA has proposed no specific limit on magnesium.

Magnesium sulfite or sulfate formation, either in the scrubber or pond, could give relatively large concentrations of sulfite and sulfate in the pond effluent water because of the high solubility of these salts. If we assume 50% oxidation of sulfite to sulfate in the scrubber, the solids to water ratio in the sluice given in the earlier tabulation, and use of a straight dolomite (no excess calcium), all of the magnesium sulfate and almost half of the magnesium sulfite formed would go into solution. The resulting concentrations would be 4,140 milligrams per liter of Mg^{++}, 4,730 milligrams of $SO_3^=$, and 12,700 milligrams of $SO_4^=$. The total salt concentration would be on the order of 22 grams per liter.

These are very high concentrations that would make release of the pond water to streams very questionable, particularly from large plants located on the smaller rivers. The large increase in sulfite concentration would be an especially severe problem. The effluent from the settling pond would be devoid of dissolved oxygen and would contain a sufficiently high residual concentration of sulfite to cause a measurable reduction of dissolved oxygen in the receiving stream. Although the oxygen reduction would not be appreciable for a 200-megawatt plant, the effluent from a much larger plant, 2,000 megawatts for example, would result in a marked effect–likely to exceed critical levels at minimum stream flows. The total oxygen reduction cannot be accurately predicted since it depends on the quantity and quality of receiving water and the rate of sulfite oxidation in such water. However, the total oxygen required to completely oxidize the sulfite from the 2,000-megawatt plant is approximately 76,000 pounds per day or a quantity sufficient, for example, to cause a dissolved oxygen reduction of over 3 milligrams per liter in the Clinch River at average flows, assuming immediate reaction and complete mixing. The reaction is not immediate and may, in fact, be quite slow; thus the actual reduction would be somewhat smaller.

As another example, the Paradise plant in the TVA system is now rated at 1,400 megawatts and will be expanded to 2,500 in 1969. The plant is located on the Green River in central Kentucky which has an average flow of about 8,000 cubic feet per second. If dolomite wet scrubbing were used on all units, discharge of the effluent would significantly reduce the dissolved oxygen content of the river and the sulfate level would increase by about 50 milligrams per liter.

The situation is quite different from that in the dry process, where little or no formation of magnesium sulfate (or sulfite) apparently takes place. The main concern in the dry process is that $MgSO_4$ will gradually form by reaction in the pond between $Mg(OH)_2$ and $CaSO_4\ 2H_2O$. The possible extent of this reaction in practice has not been determined.

If a high-calcium limestone is used, there should be many situations in which release of the dissolved calcium salts to watercourses would be acceptable. The choice then would be between the cost and difficulty of recycling and the undesirable release of even an acceptable amount of dissolved salts. If dolomitic limestone is used, the number of acceptable situations for pond water release will be drastically reduced.

Appendix C: Cost Estimates

Detailed Estimates

Table C–1. Summary of Estimated Fixed Investment Requirements: Process A – Limestone Injection-Scrubbing[a]

(200-mw existing power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	20,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	40,000
Receiving hopper, storage silo, conveyor supports and bridges, and powerhouse storage silo	50,000
Powerhouse structural steel revisions	3,000
Elevating and conveying in powerhouse	50,000
Pulverizing and air system equipment	85,000
Injection equipment (dense phase)	35,000
Two-stage wet scrubbers with pumps, structures, piping, and hold tank	500,000
Solids disposal system	
Equipment and piping	100,000
Disposal pond and land	150,000
Stack gas reheat system including supports, pumps, and piping	585,000
Central control room and equipment	85,000
Electrical and water distribution	149,000
Painting and insulation	30,000
Construction facilities	50,000
Total direct cost	1,932,000
Engineering design	193,000
Contractor fees and overhead	292,000
Contingency allowance	193,000
Total project investment	2,610,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

Table C–2. Summary of Estimated Fixed Investment Requirements: Process A – Limestone Injection-Scrubbing[a]

(1,000-mw existing power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	32,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	114,000
Receiving hopper, storage silo, conveyor supports and bridges, and powerhouse storage silo	127,000
Powerhouse structural steel revisions	6,000
Elevating and conveying in powerhouse	127,000
Pulverizing and air system equipment	285,000
Injection equipment (dense phase)	85,000
Two-stage wet scrubbers with pumps, structures, piping, and hold tank	1,630,000
Solids disposal system	
Equipment and piping	364,000
Disposal pond and land	323,000
Stack gas reheat system including supports, pumps, and piping	2,125,000
Central control room and equipment	245,000
Electrical and water distribution	463,000
Painting and insulation	120,000
Construction facilities	270,000
Total direct cost	6,316,000
Engineering design	494,000
Contractor fees and overhead	955,000
Contingency allowance	445,000
Total project investment	8,210,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

Table C–3. Summary of Estimated Fixed Investment Requirements: Process A – Limestone Injection-Scrubbing[a]

(1,000-mw new power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	20,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	100,000
Receiving hopper, storage silo, conveyor supports and bridges, and powerhouse storage silo	115,000
Elevating and conveying in powerhouse	100,000
Pulverizing and air system equipment	250,000
Injection equipment (dense phase)	85,000
Two-stage wet scrubbers with pumps, structures, piping, and hold tank	1,580,000
Solids disposal system	
Equipment and piping	362,000
Disposal pond and land	323,000
Stack gas reheat system including supports, pumps, and piping	2,125,000
Central control room and equipment	230,000
Electrical and water distribution	445,000
Painting and insulation	110,000
Construction facilities	250,000
Total direct cost	6,095,000
Engineering design	370,000
Contractor fees and overhead	745,000
Contingency allowance	410,000
Total project investment	7,620,000
Electrostatic precipitator credit[b]	(1,300,000)
Total chargeable project investment for wet-scrubbing facilities	6,320,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

[b] 99% effective electrostatic precipitator

Table C–4. Summary of Estimated Fixed Investment Requirements: Process B – Limestone Addition-Scrubbing[a]

(200-mw existing power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	25,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	45,000
Receiving hopper, storage silo, conveyor supports and bridges, surge storage, and feed conveyors	75,000
Wet grinding ball mill and classifier	115,000
Slurry storage and pumping	35,000
Two-stage wet scrubbers with pumps, piping, structures, and hold tank	505,000
Solids disposal system	
Equipment and piping	120,000
Disposal pond and land	161,000
Stack gas reheat system including supports, pumps, and piping	610,000
Central control room and equipment	85,000
Electrical and water distribution	170,000
Painting and insulation	40,000
Construction facilities	60,000
Total direct cost	2,046,000
Engineering design	205,000
Contractor fees and overhead	304,000
Contingency allowance	205,000
Total project investment	2,760,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

Table C–5. Summary of Estimated Fixed Investment Requirements: Process B – Limestone Addition-Scrubbing[a]

(1,000-mw existing power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	32,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	135,000
Receiving hopper, storage silo, conveyor supports and bridges, surge storage, and feed conveyors	190,000
Wet grinding ball mill and classifier	275,000
Slurry storage and pumping	90,000
Two-stage wet scrubbers with pumps, piping, structures, and hold tank	1,700,000
Solids disposal system	
Equipment and piping	390,000
Disposal pond and land	425,000
Stack gas reheat system including supports, pumps, and piping	2,410,000
Central control room and equipment	260,000
Electrical and water distribution	524,000
Painting and insulation	150,000
Construction facilities	300,000
Total direct cost	6,781,000
Engineering design	549,000
Contractor fees and overhead	990,000
Contingency allowance	500,000
Total project investment	8,820,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

Table C–6. Summary of Estimated Fixed Investment Requirements: Process B – Limestone Addition-Scrubbing[a]

(1,000-mw new power unit)

	Investment, $
Yard improvements	
Road and general yard modifications	25,000
Limestone storage and handling facilities	
Concrete foundations and conveyor tunnel	125,000
Receiving hopper, storage silo, conveyor supports and bridges, surge storage, and feed conveyors	170,000
Wet grinding ball mill and classifier	275,000
Slurry storage and pumping	90,000
Two-stage wet scrubbers with pumps, piping, structures, and hold tank	1,600,000
Solids disposal system	
Equipment and piping	390,000
Disposal pond and land	425,000
Stack gas reheat system including supports, pumps, and piping	2,390,000
Central control room and equipment	245,000
Electrical and water distribution	476,000
Painting and insulation	120,000
Construction facilities	260,000
Total direct cost	6,621,000
Engineering design	400,000
Contractor fees and overhead	809,000
Contingency allowance	440,000
Total project investment	8,270,000
Electrostatic precipitator credit[b]	(1,300,000)
Total chargeable project investment for wet-scrubbing facilities	6,970,000

[a] Basis:
Sulfur in coal, 3.5%
Stack gas reheat to 250° F. by indirect liquid-gas method
Direct solids disposal as 10% slurry (no return of water), scrubber-to-pond distance of 1 mi.

[b] 99% effective electrostatic precipitator

Table C–7. Annual Operating Costs for Limestone—Wet Scrubbing Power Plant Stack Gas: Process A—Limestone Injection-Scrubbing[a]

(200-mw existing unit, 2.0% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	43.2 M ton	2.05/ton	88,600	0.148
Conversion costs				
Operating labor and supervision	13,150 man-hr	4.00/man-hr	52,600	0.088
Utilities				
Water	144,000 M gal	0.10/M gal	14,400	0.024
Electricity	8,800,000 kwh	0.004/kwh	35,200	0.059
Maintenance				
Labor and material			72,500	0.121
Analyses	2,190 hr	7.50/hr	16,400	0.027
Subtotal conversion costs			191,100	0.319
Subtotal direct costs			279,700	0.467
Indirect Costs				
Capital charges, 14.5% of fixed investment			339,300	0.565
Overhead				
Plant, 20% of conversion costs			38,200	0.064
Administrative, 10% of operating labor			5,300	0.009
Subtotal indirect costs			382,800	0.629
Total annual operating cost before credits			662,500	1.096
Precipitator operating credit			−17,900	(0.030)
Thermal effect of raw limestone injection on operating cost of power generation			+11,500	0.019
Maintenance credit for corrosion reduction in boiler			−18,000	(0.030)
Total chargeable annual operating cost			638,100	1.055

[a] Basis:
Coal burned, 600,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% −200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal—scrubber to pond— 1 mile, 10% solids slurry
North Alabama plant location
Capital investment $2,340,000 fixed; $12,000 working

Table C–8. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process A–Limestone Injection-Scrubbing[a]

(200-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	77.2 M ton	2.05/ton	158,300	0.264
Conversion costs				
Operating labor and supervision	14,500 man-hr	4.00/man-hr	58,000	0.097
Utilities				
Water	252,000 M gal	0.10/M gal	25,200	0.042
Electricity	9,280,000 kwh	0.004/kwh	37,100	0.062
Maintenance				
Labor and material			78,300	0.131
Analyses	2,190 hr	7.50/hr	16,400	0.027
Subtotal conversion costs			215,000	0.359
Subtotal direct costs			373,300	0.623
Indirect Costs				
Capital charges, 14.5% of fixed investment			378,500	0.631
Overhead				
Plant, 20% of conversion costs			43,000	0.072
Administrative, 10% of operating labor			5,800	0.010
Subtotal indirect costs			427,300	0.713
Total annual operating cost before credits			800,600	1.336
Precipitator operating credit			–17,900	(0.030)
Thermal effect of raw limestone injection on operating cost of power generation			+19,400	0.032
Maintenance credit for corrosion reduction in boiler			–18,000	(0.030)
Total chargeable annual operating cost			784,100	1.308

[a] Basis:
Coal burned, 600,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $2,610,000 fixed; $20,000 working

Table C–9. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process A–Limestone Injection-Scrubbing[a]

(200-mw existing unit, 5.0% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	110.8 M ton	2.05/ton	227,100	0.378
Conversion costs				
Operating labor and supervision	15,800 man-hr	4.00/man-hr	63,200	0.105
Utilities				
Water	360,000 M gal	0.10/M gal	36,000	0.060
Electricity	11,580,000 kwh	0.004/kwh	46,300	0.077
Maintenance				
Labor and material			88,600	0.148
Analyses	2,190 hr	7.50/hr	16,400	0.027
Subtotal conversion costs			250,500	0.417
Subtotal direct costs			477,600	0.795
Indirect Costs				
Capital charges, 14.5% of fixed investment			414,700	0.691
Overhead				
Plant, 20% of conversion costs			50,100	0.083
Administrative, 10% of operating labor			6,300	0.011
Subtotal indirect costs			471,100	0.785
Total annual operating cost before credits			948,700	1.580
Precipitator operating credit			−17,900	(0.030)
Thermal effect of raw limestone injection on operating cost of power generation			+28,800	0.048
Maintenance credit for corrosion reduction in boiler			−18,000	(0.030)
Total chargeable annual operating cost			941,600	1.568

[a] Basis:
Coal burned, 600,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% −200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $2,860,000 fixed; $30,000 working

Table C–10. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process A–Limestone Injection-Scrubbing[a]

(500-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	193.2 M ton	2.05/ton	396,000	0.264
Conversion costs				
Operating labor and supervision	20,260 man-hr	4.00/man-hr	81,000	0.054
Utilities				
Water	630,000 M gal	0.10/M gal	63,000	0.042
Electricity	23,185,000 kwh	0.004/kwh	92,700	0.062
Maintenance				
Labor and material			162,700	0.108
Analyses	4,380 hr	7.50/hr	32,800	0.022
Subtotal conversion costs			432,200	0.288
Subtotal direct costs			828,200	0.552
Indirect Costs				
Capital charges, 14.5% of fixed investment			786,600	0.524
Overhead				
Plant, 20% of conversion costs			86,400	0.058
Administrative, 10% of operating labor			8,100	0.005
Subtotal indirect costs			881,100	0.587
Total annual operating cost before credits			1,709,300	1.139
Precipitator operating credit			–44,000	(0.029)
Thermal effect of raw limestone injection on operating cost of power generation			+48,500	0.032
Maintenance credit for corrosion reduction in boiler			–42,000	(0.028)
Total chargeable annual operating cost			1,671,800	1.114

[a] Basis:
Coal burned, 1,500,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $5,425,000 fixed; $40,000 working

Table C–11. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process A–Limestone Injection-Scrubbing[a]

(1,000-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	386.0 M ton	2.05/ton	791,000	0.264
Conversion costs				
Operating labor and supervision	32,740 man-hr	4.00/man-hr	131,000	0.044
Utilities				
Water	1,260,000 M gal	0.07/M gal	88,200	0.029
Electricity	45,000,000 kwh	0.003/kwh	135,000	0.045
Maintenance				
Labor and material			238,000	0.079
Analyses	6,570 hr	7.50/hr	49,300	0.016
Subtotal conversion costs			641,500	0.213
Subtotal direct costs			1,432,500	0.477
Indirect Costs				
Capital charges, 14.5% of fixed investment			1,190,500	0.396
Overhead				
Plant, 20% of conversion costs			128,300	0.043
Administrative, 10% of operating labor			13,100	0.004
Subtotal indirect costs			1,331,900	0.440
Total annual operating cost before credits			2,764,400	0.917
Precipitator operating credit			–66,000	(0.022)
Thermal effect of raw limestone injection on operating cost of power generation			+97,000	0.032
Maintenance credit for corrosion reduction in boiler			–78,000	(0.026)
Total chargeable annual operating cost			2,717,400	0.901

[a] Basis:
Coal burned, 3,000,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $8,210,000 fixed; $80,000 working

Table C–12. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process A–Limestone Injection-Scrubbing[a]

(1,000-mw new unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	386.0 M ton	2.05/ton	791,000	0.264
Conversion costs				
Operating labor and supervision	32,740 man-hr	4.00/man-hr	131,000	0.044
Utilities				
Water	1,260,000 M gal	0.07/M gal	88,200	0.029
Electricity	45,000,000 kwh	0.003/kwh	135,000	0.045
Maintenance				
Labor and material			220,000	0.073
Analyses	6,570 hr	7.50/hr	49,300	0.016
Subtotal conversion costs			623,500	0.207
Subtotal direct costs			1,414,500	0.471
Indirect Costs				
Capital charges, 13.0% of fixed investment			990,600	0.330
Overhead				
Plant, 20% of conversion costs			124,900	0.042
Administrative, 10% of operating labor			13,100	0.004
Subtotal indirect costs			1,128,600	0.376
Total annual operating cost before credits			2,543,100	0.847
Precipitator operating-investment credit			−283,600	(0.095)
Thermal effect of raw limestone injection on operating cost of power generation			+97,000	0.032
Maintenance credit for corrosion reduction in boiler			−78,000	(0.026)
Total chargeable annual operating cost			2,278,500	0.758

[a] Basis:
Coal burned, 3,000,000 tons/yr
99% effective electrostatic precipitator replaced by two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 110% stoichiometric calcium oxide
Ground limestone injected, 75% −200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $7,620,000 fixed; $80,000 working

Table C–13. Annual Operating Costs for Limestone—Wet Scrubbing Power Plant Stack Gas: Process B–Limestone Addition-Scrubbing[a]

(200-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	93.0 M ton	2.05/ton	191,100	0.319
Conversion costs				
Operating labor and supervision	15,910 man-hr	4.00/man-hr	63,600	0.106
Utilities				
Water	285,000 M gal	0.10/M gal	28,500	0.047
Electricity	9,600,000 kwh	0.004/kwh	38,400	0.064
Maintenance				
Labor and material			85,600	0.143
Analyses	2,190 hr	7.50/hr	16,400	0.027
Subtotal conversion costs			232,500	0.387
Subtotal direct costs			423,600	0.706
Indirect Costs				
Capital charges, 14.5% of fixed investment			400,200	0.667
Overhead				
Plant, 20% of conversion costs			46,500	0.077
Administrative, 10% of operating labor			6,400	0.011
Subtotal indirect costs			453,100	0.745
Total annual operating cost before credits			876,700	1.451
Precipitator operating credit			–17,900	(0.030)
Total chargeable annual operating cost			858,800	1.421

[a] Basis:
Coal burned, 600,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 130% stoichiometric calcium oxide
Ground limestone added, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $2,760,000 fixed; $25,000 working

Table C–14. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process B–Limestone Addition-Scrubbing[a]

(500-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	233.2 M ton	2.05/ton	478,000	0.319
Conversion costs				
Operating labor and supervision	24,670 man-hr	4.00/man-hr	98,700	0.066
Utilities				
Water	713,000 M gal	0.10/M gal	71,300	0.048
Electricity	23,500,000 kwh	0.004/kwh	94,000	0.063
Maintenance				
Labor and material			180,700	0.120
Analyses	4,380 hr	7.50/hr	32,800	0.022
Subtotal conversion costs			477,500	0.319
Subtotal direct costs			955,500	0.635
Indirect Costs				
Capital charges, 14.5% of fixed investment			845,400	0.564
Overhead				
Plant, 20% of conversion costs			95,500	0.064
Administrative, 10% of operating labor			9,900	0.007
Subtotal indirect costs			950,800	0.635
Total annual operating cost before credits			1,906,300	1.270
Precipitator operating credit			–44,000	(0.029)
Total chargeable annual operating cost			1,862,300	1.241

[a] Basis:
Coal burned, 1,500,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 130% stoichiometric calcium oxide
Ground limestone added, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $5,830,000 fixed; $50,000 working

Table C—15. Annual Operating Costs for Limestone—Wet Scrubbing Power Plant Stack Gas: Process B—Limestone Addition-Scrubbing[a]

(1,000-mw existing unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	466.4 M ton	2.05/ton	956,000	0.319
Conversion costs				
Operating labor and supervision	33,430 man-hr	4.00/man-hr	133,700	0.044
Utilities				
Water	1,426,000 M gal	0.07/M gal	99,800	0.033
Electricity	45,666,000 kwh	0.003/kwh	137,000	0.046
Maintenance				
Labor and materials			255,800	0.085
Analyses	6,570 hr	7.50/hr	49,300	0.016
Subtotal conversion costs			675,600	0.224
Subtotal direct costs			1,631,600	0.543
Indirect Costs				
Capital charges, 14.5% of fixed investment			1,278,900	0.426
Overhead				
Plant, 20% of conversion costs			135,100	0.045
Administrative, 10% of operating labor			13,400	0.004
Subtotal indirect costs			1,427,400	0.465
Total annual operating cost before credits			3,059,000	1.008
Precipitator operating credit			−66,000	(0.022)
Total chargeable annual operating cost			2,993,000	0.986

[a] Basis:
Coal burned, 3,000,000 tons/yr
90% effective electrostatic precipitator not in operation, two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 130% stoichiometric calcium oxide
Ground limestone added, 75% −200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal—scrubber to pond—1 mile, 10% solids slurry
North Alabama plant location
Capital investment $8,820,000 fixed; $96,000 working

Table C–16. Annual Operating Costs for Limestone–Wet Scrubbing Power Plant Stack Gas: Process B–Limestone Addition-Scrubbing[a]

(1,000-mw new unit, 3.5% S in coal)

	Annual quantity	$/unit	Total annual cost, $	$/ton of coal burned
Direct Costs				
Delivered raw material				
Limestone, 95% CO_3	466.4 M ton	2.05/ton	956,000	0.319
Conversion costs				
Operating labor and supervision	33,430 man-hr	4.00/man-hr	133,700	0.044
Utilities				
Water	1,426,000 M gal	0.07/M gal	99,800	0.033
Electricity	45,666,000 kwh	0.003/kwh	137,000	0.046
Maintenance				
Labor and material			239,800	0.080
Analyses	6,570 hr	7.50/hr	49,300	0.016
Subtotal conversion costs			659,600	0.219
Subtotal direct costs			1,615,600	0.538
Indirect Costs				
Capital charges, 13.0% of fixed investment			1,075,100	0.358
Overhead				
Plant, 20% of conversion costs			131,900	0.044
Administrative, 10% of operating labor			13,400	0.004
Subtotal indirect costs			1,220,400	0.406
Total annual operating cost before credits			2,836,000	0.944
Precipitator operating-investment credit			–283,600	(0.095)
Total chargeable annual operating cost			2,552,400	0.849

[a] Basis:

Coal burned, 3,000,000 tons/yr
99% effective electrostatic precipitator replaced by two-stage scrubbing
Stack gas reheat from 118° to 250° F., indirect liquid-gas method
Limestone added, 130% stoichiometric calcium oxide
Ground limestone added, 75% –200 mesh
Coal, 0.75 lb/kwh
Unit on-stream time, 8,000 hr
Direct solids disposal–scrubber to pond–1 mile, 10% solids slurry
North Alabama plant location
Capital investment $8,270,000 fixed; $96,000 working

Determination of Credit for Corrosion Reduction

Injection of limestone into the boiler is expected to reduce sulfur trioxide concentration in the exhaust gases to a very low value. Since this will greatly reduce the formation of acidic salts at high temperature and sulfuric acid at low temperature, corrosion of the boiler heat transfer surfaces will be reduced. The expected corrosion reduction will result in a cost credit to be applied in the evaluation of process A.

As indicated by TVA experience and the Federal Power Commission report[1] the average maintenance cost for a boiler is $0.45/ton of coal burned. From TVA accounts of maintenance distribution, it was estimated that expected corrosion reduction could be valued at $0.11/ton of coal burned. However, corrosion reduction must be weighed against increased surface erosion due to the limestone. A rough estimate of this offsetting effect indicates that no more than about 30% of the above credit could be saved. Therefore, a credit of $0.03/ton of coal burned was applied in evaluating process A.

Determination of Operating and Investment Credits for Eliminating Electrostatic Precipitator

In outlining the basis for evaluation of limestone-wet scrubbing processes, it was assumed that existing fossil fuel-burning units are equipped with 90% effective electrostatic precipitators and that new units would be equipped with at least 99% effective precipitators if no other means of stack gas cleaning were to be used. If wet scrubbers are added to the system for better sulfur dioxide removal, the need for the electrostatic precipitators becomes questionable. In the section "Economic Evaluation" an economic appraisal of existing units using various precipitator-scrubber combinations indicated that it would be economical to shut down the existing precipitators and remove the dust with the scrubbers alone. Of course, the installation of electrostatic precipitators on new units would not be required if a wet-scrubbing process were used.

For the use of processes A and B on existing units, an operating credit was taken; for new units, both an operating and investment credit were taken. The estimated credits are shown below:

Precipitator Annual Credits

	Existing[a]		New[b]
	200 mw	1,000 mw	1,000 mw
Labor and overheads	$ 3,600	$10,000	$ 13,000
Power	10,700	38,000	75,600
Maintenance	3,600	18,000	26,000
Total credit for existing precipitators	17,900	66,000	–
Capital charges at 13% of new investment[c]	–	–	169,000
Total credit for new precipitator	–	–	$283,600

[a] 90% efficient.
[b] 99% efficient.
[c] 1,000-mw new unit; 3,300,000 actual c.f.m.; 3.5% S in coal; estimated project cost of 99% effective precipitator–$1,300,000.

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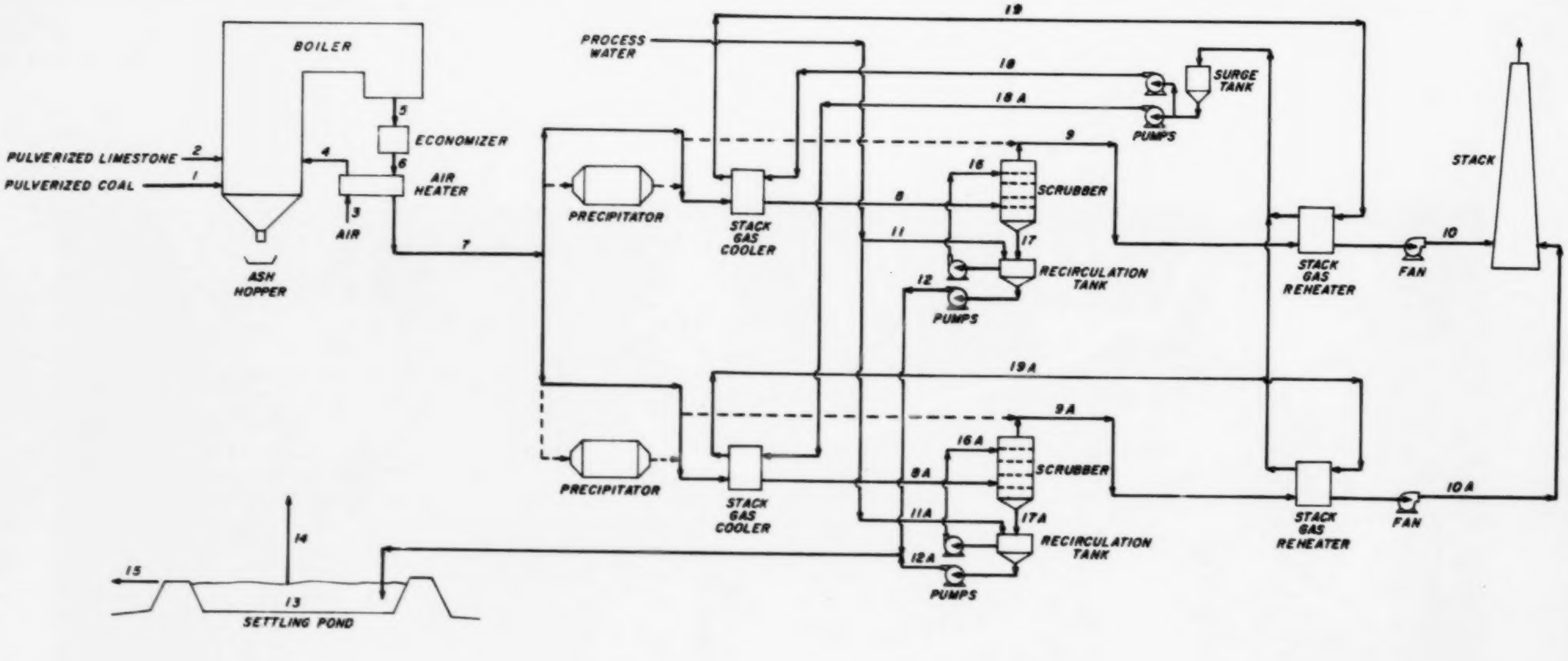
[1] Federal Power Commission. *Steam-Electric Plant Construction Cost and Annual Production Expenses*, Nineteenth Annual Supplement (1966).

Appendix D: Drawings


PULVERIZED LIMESTONE
PULVERIZED COAL
BOILER
ASH HOPPER
ECONOMIZER
AIR HEATER
AIR
PROCESS WATER
PRECIPITATOR
STACK GAS COOLER
SCRUBBER
RECIRCULATION TANK
PUMPS
SURGE TANK
STACK GAS REHEATER
FAN
STACK
SETTLING POND
1
2
3
4
5
6
7
8
8A
9
9A
10
10A
11
11A
12
12A
13
14
15
16
16A
17
17A
18
18A
19
19A

TYPICAL OPERATING CONDITIONS FOR LIME SCRUBBING PROCESS WITH A 200 MEGAWATT STEAM PLANT - 3.5 % SULFUR IN COAL

STREAM NO.	1	2	3	4	5	6	7	8 & 8A	9 & 9A	10 & 10A
DESCRIPTION	COAL TO BOILER	LIMESTONE TO BOILER	COMBUSTION AIR TO AIR HEATER	COMBUSTION AIR TO BOILER	GAS TO ECONOMIZER	GAS TO AIR HEATER	GAS TO COOLERS	GAS TO SCRUBBER	GAS TO REHEATER	GAS TO STACK
RATE, LBS./HR.	150M	19.3M	1,911M	1,720M	1,869M	1,869M	2,096M	1,028M	1,045M	1,045M
SCFM	-	-	419M	379M	400M	400M	440M	220M	227M	227M
GPM	-	-	-	-	-	-	-	-	-	-
PARTICULATES, LBS./HR.	-	-	0	0	26.6M	26.6M	26.6M	13.3M	0.17M	0.17M
TEMPERATURE, °F	AMBIENT	AMBIENT	110	610	690	705	310	177	120	280
SPECIFIC GRAVITY										
VISCOSITY, CPS			0.019	0.030						
UNDISSOLVED SOLIDS, %										
pH										

STREAM NO.	11 & 11A	12 & 12A	13	14	15	16 & 16A	17 & 17A	18 & 18A	19 & 19A
DESCRIPTION	MAKE-UP WATER	SLURRY TO POND	POND SOLIDS ACCUMULATION	POND WATER LOSS	POND OVERFLOW WATER	RECYCLE SLURRY TO SCRUBBER	SCRUBBER SLURRY TO RECIR. TANK	WARM WATER TO STACK GAS COOLER	HOT WATER TO STACK GAS REHEATER
RATE, LBS./HR.	200M	196M	39M	124M	226M	4,200M	4,196M	273M	273M
SCFM	-	-	-	-	-	-	-	-	-
GPM	400	373	-	249	456	8M	7,992	556	587
PARTICULATES, LBS./HR.	-	-	-	-	40	-	-	-	-
TEMPERATURE, °F	65	120		65	65	120	120	148	280
SPECIFIC GRAVITY	1.0	1.05			1.0	1.05	1.05	0.98	0.93
VISCOSITY, CPS	1.05				1.05			0.44	0.20
UNDISSOLVED SOLIDS, %	0	10				10	10.5		
pH	7								

CALCULATED ANALYSIS OF GAS FROM AIR HEATER

	CONCENTRATION (VOLUME %)			PARTICULATES (GRAMS / SCF)		
	% S IN COAL			% S IN COAL		
	2.0	3.5	5.0	2.0	3.5	5.0
SULFUR DIOXIDE	0.09	0.16	0.23	-	-	-
WATER VAPOR	7.84	7.89	7.94	-	-	-
NITROGEN	74.53	74.36	74.18	-	-	-
OXYGEN	4.86	4.84	4.82	-	-	-
CARBON DIOXIDE	12.68	12.75	12.83	-	-	-
FLY ASH	-	-	-	3.72	3.80	3.90
CaO & CaSO$_4$	-	-	-	1.83	3.25	4.65

NOTES:

1. CALCULATIONS BASED ON:
   - a. 110% STOICHIOMETRIC LIME
   - b. % SULFUR IN COAL (DRY BASIS)
   - c. 12 % ASH COAL (AS FIRED BASIS)
   - d. 92 % OF SULFUR IN COAL EVOLVES AS SO$_2$
   - e. 75 % OF ASH IN COAL EVOLVES AS FLY ASH
   - f. MISCELLANEOUS INFORMATION FROM SCRUBBER MANUFACTURERS
   - g. 98.7 % REMOVAL OF PARTICULATES TO SCRUBBERS
   - h. 95 % TOTAL SO$_2$ REMOVAL
   - i. LIMESTONE TO BOILER CONTAINS 10 % WATER AND 95% CaCO$_3$ (DRY BASIS).
2. POND WATER LOSS CALCULATIONS: EVAPORATION FROM U.S.D.A. TECH. BULLETIN 271; ASSUME 32 IN./YR. RAINFALL; ASSUME 1×10^{-6} CM / SEC. SEEPAGE.
3. PARTICULATE RATES IN LBS./HR. SHOULD BE ADDED TO STACK GAS RATES TO GET TOTAL STREAM RATE (e.g. TO DETERMINE THE TOTAL RATE OF STREAM 5, ADD 1,869M & 26.6M)
4. STREAMS SUCH AS 8 & 8A, 9 & 9A HAVE EQUAL RATES & PHYSICAL PROPERTIES, BUT ONLY THE INFORMATION FOR ONE SCRUBBER SYSTEM IS LISTED IN THE TABLE.
5. GAS DUCT BYPASSES ARE SHOWN AS DASHED LINES.
6. PRECIPITATORS ARE TO BE USED ONLY WHEN SCRUBBERS ARE SHUTDOWN AND BYPASSED.

SYMBOL IN TABLE

M----- THOUSAND

Figure D–1. Flowsheet – Process A

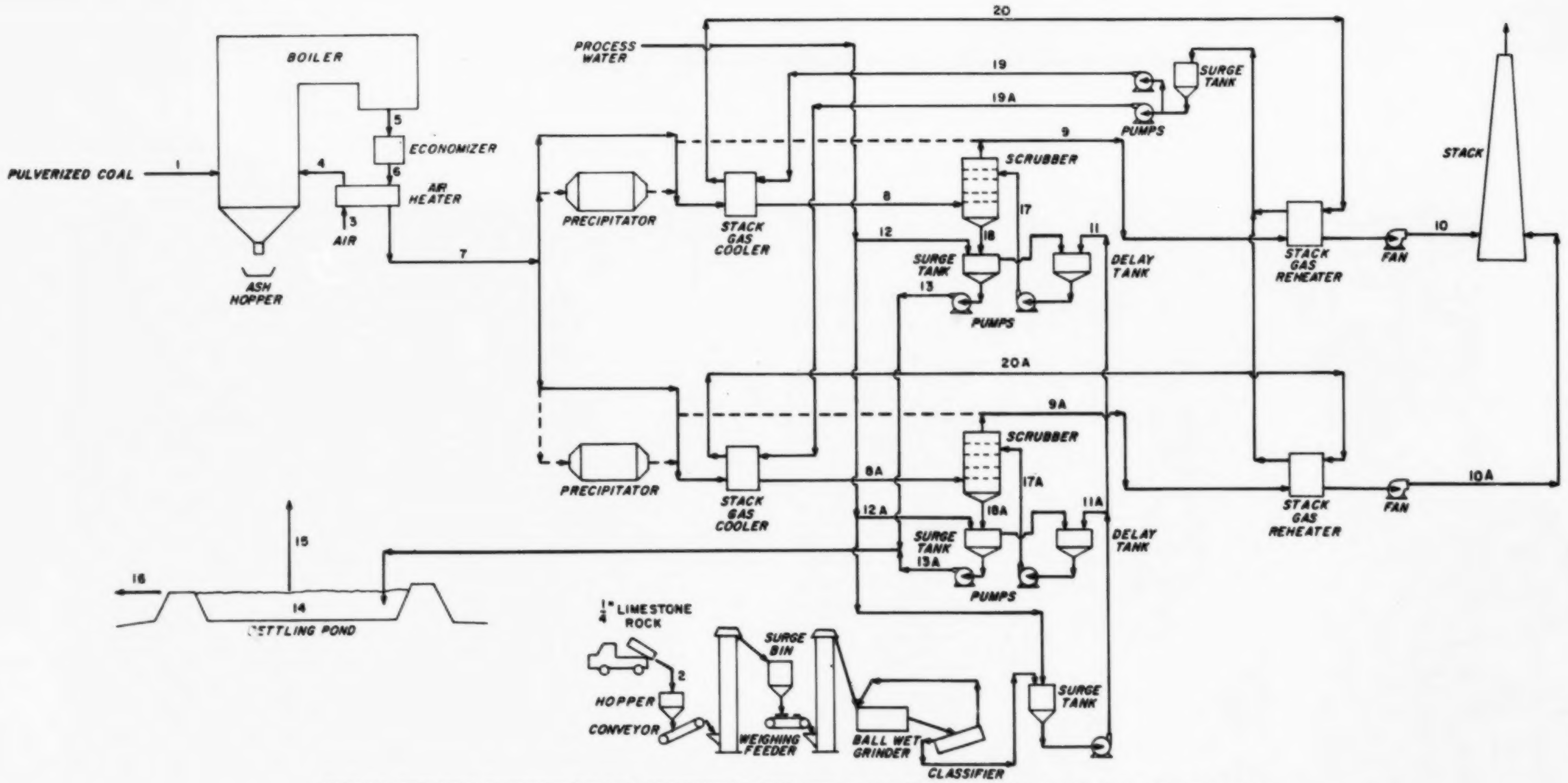
TYPICAL OPERATING CONDITIONS FOR LIME SCRUBBING PROCESS WITH A 20C MEGAWATT STEAM PLANT - 2 % SULFUR IN COAL

STREAM NO.	1	2	3	4	5	6	7	8&8A	9&9A	10&10A	11&11A	12&12A	13	14	15	16&16A	17&17A	18&18A	19&19A	
DESCRIPTION	COAL TO BOILER	LIMESTONE TO BOILER	COMBUSTION AIR TO AIR HEATER	COMBUSTION AIR TO BOILER	GAS TO ECONOMIZER	GAS TO AIR HEATER	GAS TO COOLERS	GAS TO SCRUBBER	GAS TO REHEATER	GAS TO STACK	MAKE-UP WATER	SLURRY TO POND	POND SOLIDS ACCUMULATION	POND WATER LOSS	POND OVERFLOW WATER	RECYCLE SLURRY TO SCRUBBER	SCRUBBER SLURRY TO RECIR. TANK	WARM WATER TO STACK GAS COOLER	HOT WATER TO STACK GAS REHEATER	
RATE, LBS./HR.	150M	10.8M	1,911M	1,724M	1,862M	1,862M	2,049M	1,024M	1,041M	1,041M	145M	139M	27.8M	91M	159M	4,200M	4,194M	271M	271M	
SCFM	—	—	419M	378M	399M	399M	439M	219M	226M	226M	—	—	—	—	—	—	—	—	—	
GPM	—	—	—	—	—	—	—	—	—	—	290	265	—	182	318	8M	7,989	552	583	
PARTICULATES, LBS./HR.	—	—	0	0	20.9M	20.9M	20.9M	10.4M	0.13M	0.13M	—	—	—	—	30	—	—	—	—	
TEMPERATURE, °F.	AMBIENT	AMBIENT	110	610	890	705	310	177	119	250	65	119		65	65	119	119	148	280	
SPECIFIC GRAVITY											1.0	1.05			1.0	1.05	1.05	0.98	0.93	
VISCOSITY, CPS			0.019	0.030							1.05				1.05			0.44	0.20	
UNDISSOLVED SOLIDS, %											0	10				10	10.3			
pH											7									

TYPICAL OPERATING CONDITIONS FOR LIME SCRUBBING PROCESS WITH A 200 MEGAWATT STEAM PLANT- 5 % SULFUR IN COAL

STREAM NO.	1	2	3	4	5	6	7	8&8A	9&9A	10&10A	11&11A	12&12A	13	14	15	16&16A	17&17A	18&18A	19&19A	
DESCRIPTION	COAL TO BOILER	LIMESTONE TO BOILER	COMBUSTION AIR TO AIR HEATER	COMBUSTION AIR TO BOILER	GAS TO ECONOMIZER	GAS TO AIR HEATER	GAS TO COOLERS	GAS TO SCRUBBER	GAS TO REHEATER	GAS TO STACK	MAKE-UP WATER	SLURRY TO POND	POND SOLIDS ACCUMULATION	POND WATER LOSS	POND OVERFLOW WATER	RECYCLE SLURRY TO SCRUBBER	SCRUBBER SLURRY TO RECIR. TANK	WARM WATER TO STACK GAS COOLER	HOT WATER TO STACK GAS REHEATER	
RATE, LBS./HR	150M	27.7M	1,911M	1,728M	1,876M	1,876M	2,063M	1,031M	1,052M	1,052M	255M	251M	50.2M	124M	328M	4,200M	4,196M	275M	275M	
SCFM	—	—	419M	378M	401M	401M	441M	220M	229M	229M	—	—	—	—	—	—	—	—	—	
GPM	—	—	—	—	—	—	—	—	—	—	510	478	—	248	656	8M	7,992	560	591	
PARTICULATES, LBS./HR.	—	—	0	0	32.3M	32.3M	32.3M	16.1M	0.21M	0.21M	—	—	—	—	60	—	—	—	—	
TEMPERATURE, °F	AMBIENT	AMBIENT	110	610	890	705	310	179	122	250	65	122		65	65	122	122	150	280	
SPECIFIC GRAVITY											1.0	1.05			1.0	1.05	1.05	0.98	0.93	
VISCOSITY, CPS			0.019	0.030							1.05				1.05			0.45	0.20	
UNDISSOLVED SOLIDS, %											0	10				10	10.6			
pH											7									

Figure D–2. Supplement to Figure D–1.


PULVERIZED COAL
1
BOILER
ASH HOPPER
4
5
ECONOMIZER
6
AIR HEATER
3
AIR
7
PRECIPITATOR
PRECIPITATOR
PROCESS WATER
STACK GAS COOLER
STACK GAS COOLER
20
19
19A
9
SURGE TANK
PUMPS
SCRUBBER
8
12
17
18
SURGE TANK
13
PUMPS
11
DELAY TANK
20A
9A
SCRUBBER
8A
12 A
17A
18A
SURGE TANK
13A
PUMPS
11A
DELAY TANK
STACK GAS REHEATER
STACK GAS REHEATER
FAN
FAN
10
10A
STACK
16
15
14
SETTLING POND
1/4" LIMESTONE ROCK
2
HOPPER
CONVEYOR
SURGE BIN
WEIGHING FEEDER
BALL WET GRINDER
CLASSIFIER
SURGE TANK

TYPICAL OPERATING CONDITIONS FOR LIME SCRUBBING PROCESS WITH A 200 MEGAWATT STEAM PLANT - 3.5 % SULFUR IN COAL
2-STAGE SCRUBBERS, 85% SO_2 REMOVAL

STREAM NO.	1	2	3	4	5	6	7	8 & 8A	9 & 9A	10 & 10A	11 & 11A	12 & 12A	13 & 13A	14	15	16	17 & 17A	18 & 18A	19 & 19A	20 & 20A
	COAL TO BOILER	LIMESTONE TO HOPPER	COMBUSTION AIR TO AIR HEATER	COMBUSTION AIR TO BOILER	GAS TO ECONOMIZER	GAS TO AIR HEATER	GAS TO COOLERS	GAS TO SCRUBBER	GAS TO REHEATER	GAS TO STACK	LIMESTONE SLURRY TO DELAY TANK	MAKE-UP WATER	SLURRY TO POND	POND SOLIDS ACCUMULATION	POND WATER LOSS	POND OVERFLOW WATER	RECYCLE SLURRY TO SCRUBBER	SCRUBBER SLURRY TO SURGE TANK	WARM WATER TO STACK GAS COOLER	HOT WATER TO STACK REHEATER
RATE, LBS./HR	150 M	23.3 M	1,911 M	1,724 M	1,863 M	1,863 M	2,050 M	1,025 M	1,042 M	1,042 M	27.8 M	189 M	208 M	414 M	124 M	250 M	5,593 M	5,584 M	271 M	271 M
SCFM	-	-	419 M	378 M	399 M	399 M	439 M	219 M	225 M	225 M	-	-	-	-	-	-	-	-	-	-
GPM	-	-	-	-	-	-	-	-	-	-	44.5	378	396	-	248	500	10.7 M	10.6 M	551	583
PARTICULATES, LBS./HR.	-	-	0	0	13.5 M	13.5 M	13.5 M	6.7 M	90	90	-	-	-	-	-	50	-	-	-	-
TEMPERATURE, °F	AMBIENT	AMBIENT	110	610	890	705	310	175	118	250		65	118		65	65	118	118	146	280
SPECIFIC GRAVITY											1.25	1.0	1.05			1.0	1.05	1.05	0.98	0.93
VISCOSITY, CPS			0.019	0.030								1.05				1.05			0.45	0.20
UNDISSOLVED SOLIDS, %											35	0	10				10.1	10.4		
pH												7								

CALCULATED ANALYSIS OF GAS FROM AIR HEATER

	CONCENTRATION (VOLUME %)	PARTICULATES (GRAINS / SCF)
SULFUR DIOXIDE	0.22	
WATER VAPOR	7.76	
NITROGEN	74.58	
OXYGEN	4.89	
CARBON DIOXIDE	12.55	
FLY ASH		3.59

NOTES:

1. CALCULATIONS BASED ON:
   a. 130 % STOICHIOMETRIC LIME
   b. % SULFUR IN COAL (DRY BASIS)
   c. 12 % ASH COAL (AS FIRED BASIS)
   d. 92 % OF SULFUR IN COAL EVOLVES AS SO_2
   e. 75 % OF ASH IN COAL EVOLVES AS FLY ASH
   f. MISCELLANEOUS INFORMATION FROM SCRUBBER MANUFACTURERS
   g. 90 % REMOVAL OF FLY ASH IN PRECIPITATORS WHEN USED
   h. % REMOVAL OF PARTICULATES IN SCRUBBERS: 98.7
   i. % TOTAL SO_2 REMOVAL: SEE ABOVE
   j. LIMESTONE TO SYSTEM CONTAINS 10 % WATER AND 95 % $CaCO_3$ (DRY BASIS). ASSUME 2% LOSS.
2. POND WATER LOSS CALCULATIONS: EVAPORATION FROM U.S.D.A. TECH. BULLETIN 271; ASSUME 52 IN./YR. RAINFALL ASSUME 1×10^{-6} CM / SEC. SEEPAGE
3. PARTICULATE RATES IN LBS./HR. SHOULD BE ADDED TO STACK GAS RATES TO GET TOTAL STREAM RATE (e.g. TO DETERMINE THE TOTAL RATE OF STREAM 5, ADD 1863M & 13.5 M)
4. STREAMS SUCH AS 8 & 8A, 9 & 9A HAVE EQUAL RATES & PHYSICAL PROPERTIES, BUT ONLY THE INFORMATION FOR ONE SCRUBBER SYSTEM IS LISTED IN THE TABLE.
5. GAS DUCT BYPASSES ARE SHOWN AS DASHED LINES

SYMBOL IN TABLE

M----- THOUSAND

Figure D–3. Flowsheet – Process B

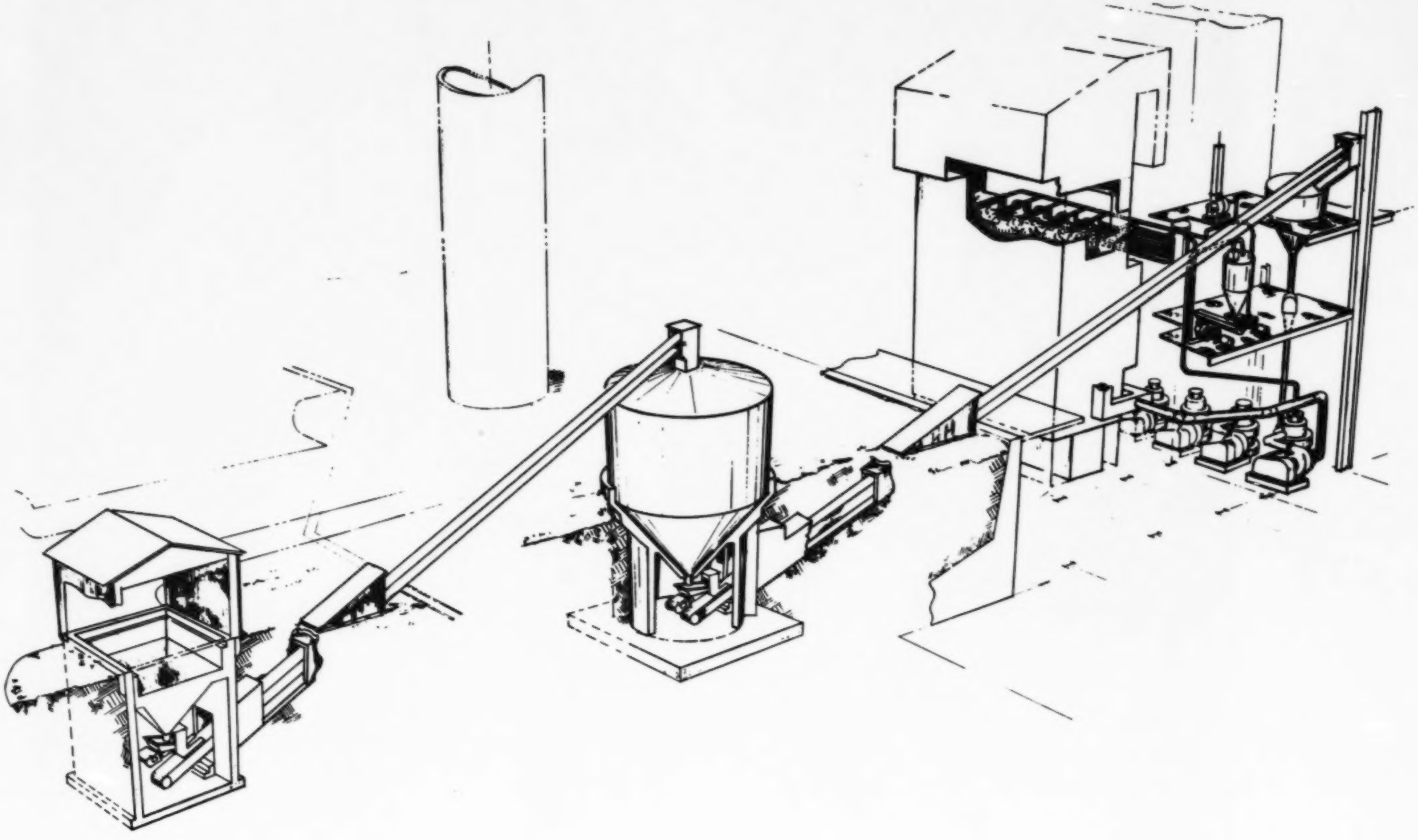


Figure D—4. Limestone receiving, handling, and dry injection facilities

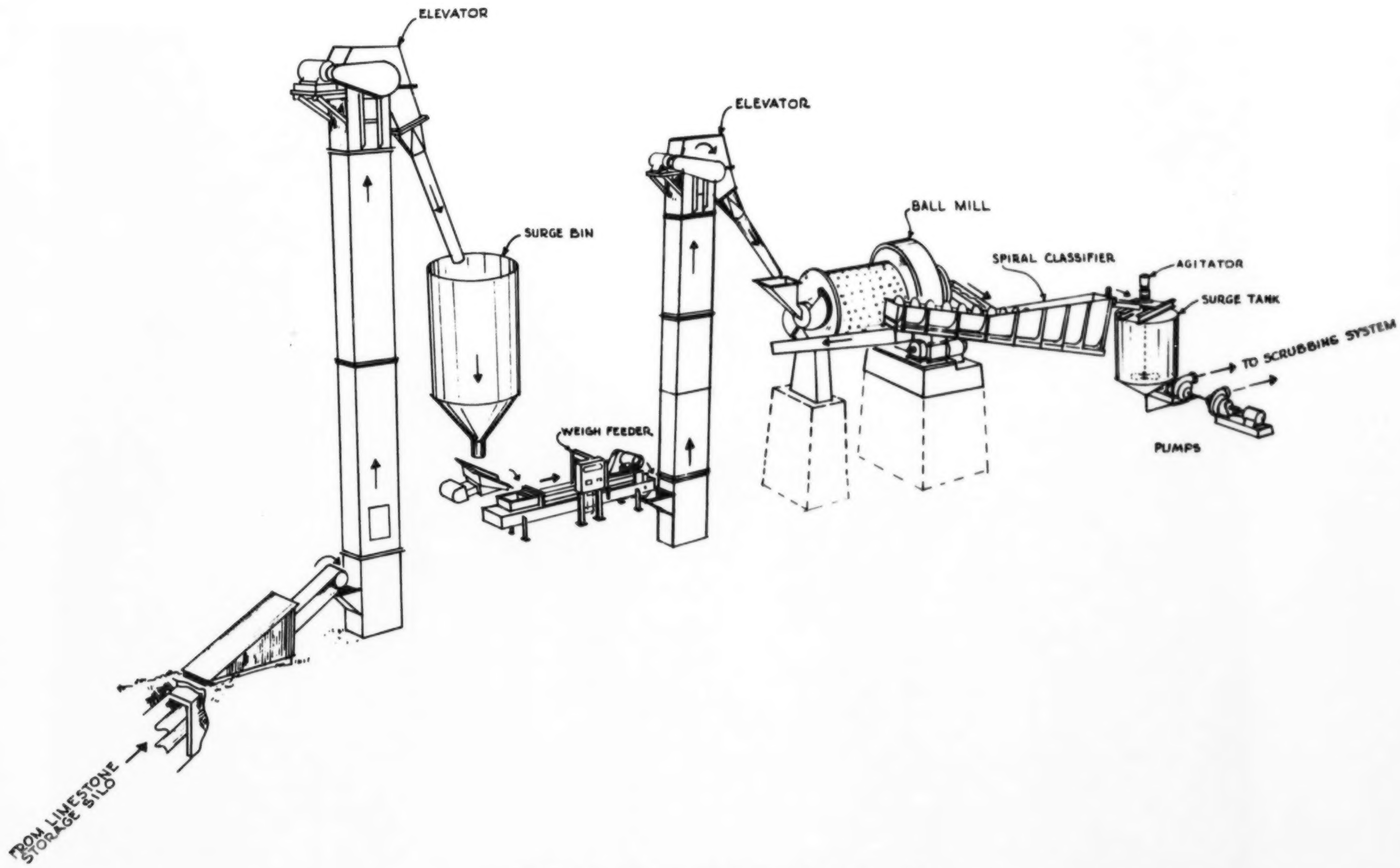

ELEVATOR
SURGE BIN
WEIGH FEEDER
ELEVATOR
BALL MILL
SPIRAL CLASSIFIER
AGITATOR
SURGE TANK
TO SCRUBBING SYSTEM
PUMPS
FROM LIMESTONE STORAGE SILO

Figure D–5. Limestone wet-grinding facilities – Process B

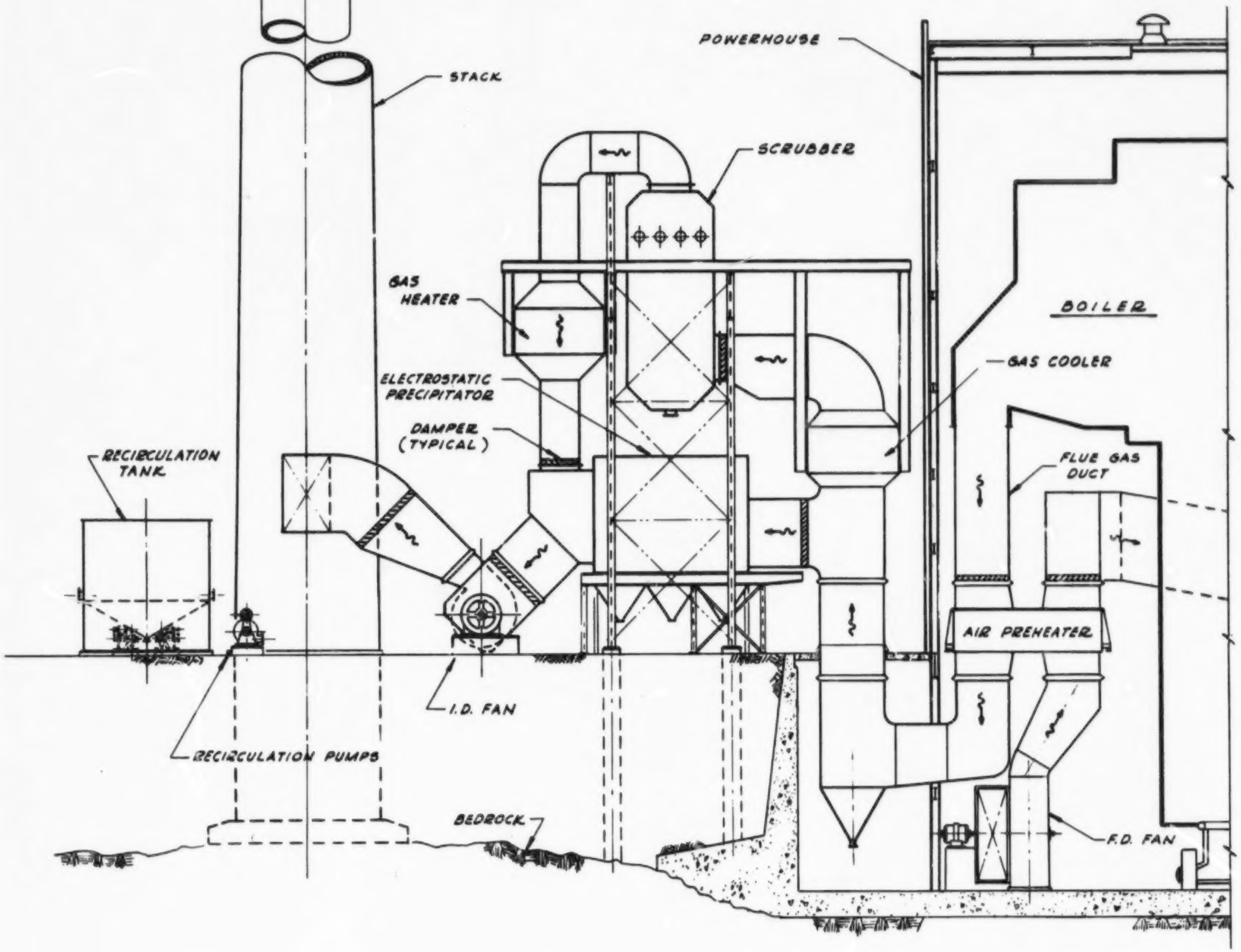

STACK
SCRUBBER
POWERHOUSE
GAS HEATER
BOILER
GAS COOLER
ELECTROSTATIC PRECIPITATOR
DAMPER (TYPICAL)
RECIRCULATION TANK
FLUE GAS DUCT
AIR PREHEATER
I.D. FAN
RECIRCULATION PUMPS
BEDROCK
F.D. FAN

Figure D–6. General arrangement–scrubber area elevation view

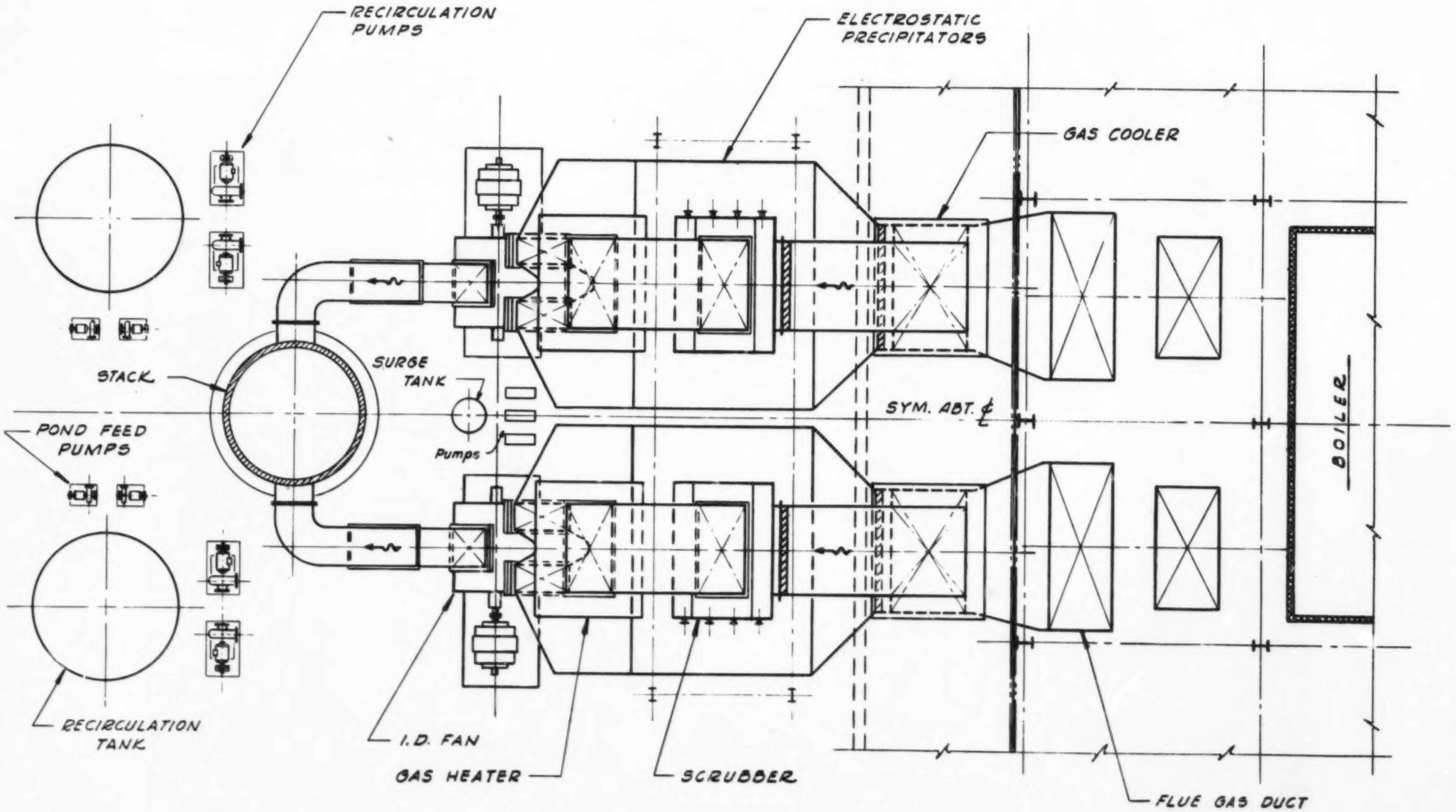

RECIRCULATION PUMPS
ELECTROSTATIC PRECIPITATORS
GAS COOLER
STACK
SURGE TANK
SYM. ABT. ℄
BOILER
POND FEED PUMPS
Pumps
RECIRCULATION TANK
I.D. FAN
GAS HEATER
SCRUBBER
FLUE GAS DUCT

Figure D–7. General arrangement–scrubber area plan view

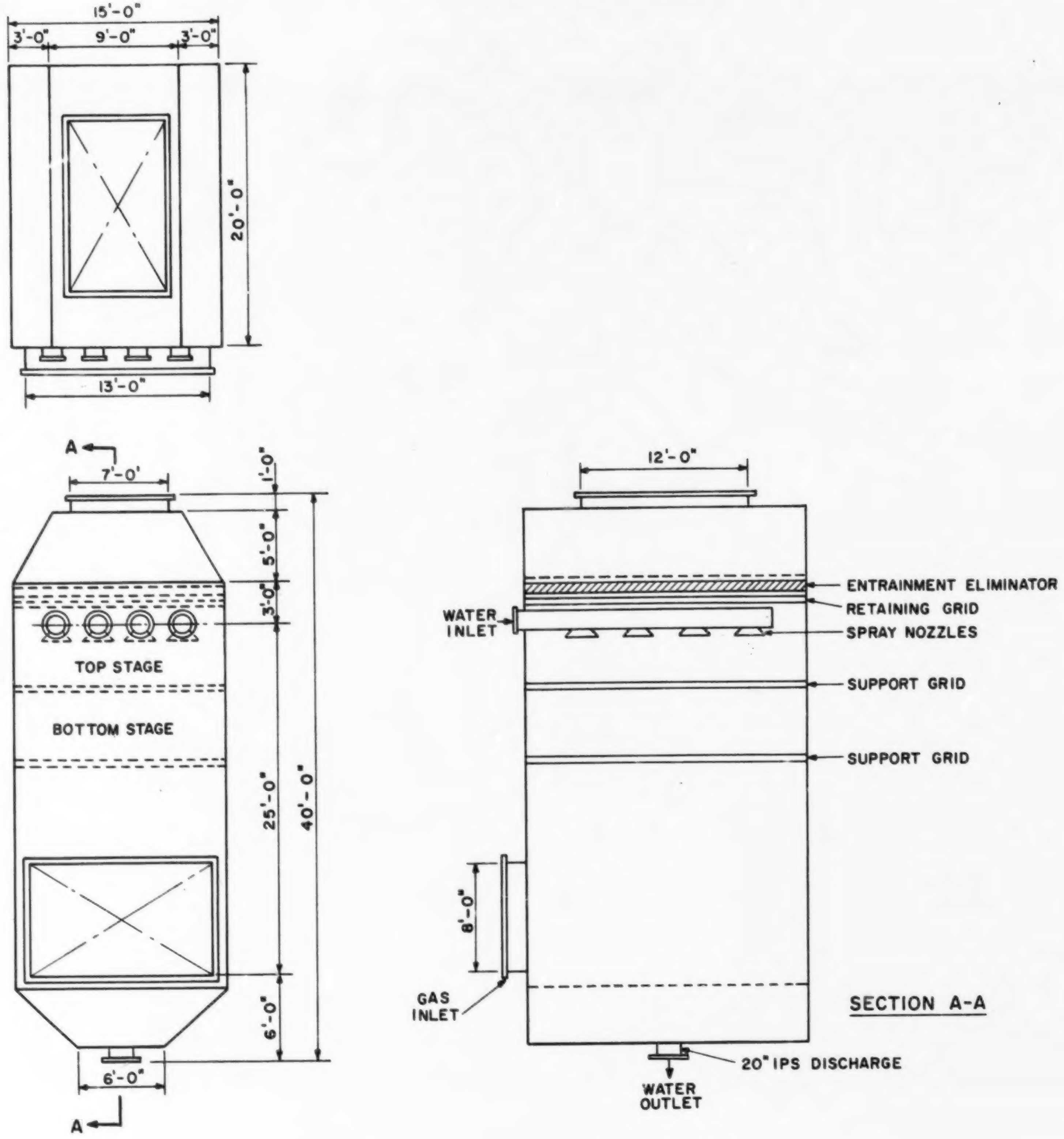

15'-0"
3'-0"
9'-0"
3'-0"
20'-0"
13'-0"
A
7'-0'
1'-0"
5'-0"
3'-0"
TOP STAGE
BOTTOM STAGE
25'-0"
40'-0"
6'-0"
6'-0"
A
12'-0"
WATER INLET
ENTRAINMENT ELIMINATOR
RETAINING GRID
SPRAY NOZZLES
SUPPORT GRID
SUPPORT GRID
8'-0"
GAS INLET
SECTION A-A
20" IPS DISCHARGE
WATER OUTLET

Figure D–8 TCA scrubber for wet limestone scrubbing process